SIMULATION AND INTEGRATION OF LIQUEFIED NATURAL GAS (LNG) PROCESSES

A Thesis

by

SAAD ALI AL-SOBHI

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

December 2007

Major Subject: Chemical Engineering

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Approved by:

Chair of Committee,	Mahmoud El-Halwagi
Committee Members,	John Baldwin
	Eyad Masad
Head of Department,	Michael Pishko

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ABSTRACT

Simulation and Integration of Liquefied Natural Gas

(LNG) Processes. (December 2007)

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The global use of natural gas is growing quickly. This is primarily attributed to its favorable characteristics and to the environmental advantages it enjoys over other fossil fuels such as oil and coal. One of the key challenges in supplying natural gas is the form (phase) at which it should be delivered. Natural gas may be supplied to the consumers as a compressed gas through pipelines. Another common form is to be compressed, refrigerated and supplied as a liquid known as *liquefied natural gas (LNG)*. When there is a considerable distance involved in transporting natural gas, LNG is becoming the preferred method of supply because of technical, economic, and political reasons. Thus, LNG is expected to play a major role in meeting the global energy demands.

This work addresses the simulation and optimization of an LNG plant. First, the process flowsheet is constructed based on a common process configuration. Then, the key units are simulated using ASPEN Plus to determine the characteristics of the various pieces of equipment and streams in the plant. Next, process integration techniques are used to optimize the process. Particular emphasis is given to energy objectives through three activities. First, the synthesis and retrofitting of a heat-exchange network are considered to reduce heating and cooling utilities. Second, the turbo-expander system is analyzed to reduce the refrigeration consumption in the process. Third, the process cogeneration is introduced to optimize the combined heat and power of the plant.

These activities are carried out using a combination of graphical, computer-aided, and mathematical programming techniques. A case study on typical LNG facilities is solved to examine the benefits of simulation and integration of the process. The technical, economic, and environmental impact of the process modifications are also discussed.

DEDICATION

TO

MY PARENTS

MY WIFE

MY PROSPECTIVE BABY

MY BROTHERS AND SISTERS

WITH LOVE

ACKNOWLEDGEMENTS

All the praises are due to Allah, the Most Beneficent and the Most Merciful for blessing me with the ability to pursue my graduate studies and seek knowledge.

There are many people who have helped me in this reserch. I would like to express my gratitude to my advisor, Dr. Mahmoud El-Halwagi, for his guidance throughout this research. I have been honored by working with Dr. El-Halwagi. His endless help has been one of the most important key factors that always motivate me to work and do my research. Also, I am thankful to all of my graduate and undergraduate studies instructors. Special thanks are due to Dr. Hassan Al-Fadala, my academic co-advisor, who introduced me to Dr. El-Halwagi and Process Integration field. I am personally indebted to Dr. Al-Fadala for his advises and concerns.

I would like to extend my appreciation to my committee members, Dr. John Baldwin and Dr. Eyad Masad, for serving in my research committee and for their comments and valuable time.

Thanks are extended to all of my colleagues, the Process Integration Group members, for their support and help. Special thanks to Abdullah, Musaed, Nasser, and my officemates Eid, Ahmed, Jose, and Viet.

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1. INTRODUCTION

1.1 Natural Gas: Resources and Reserves

Natural gas is a vital commodity in the global energy market. The current status of primary sources of energy is summarized in Figure 1.1. Clearly, oil is the leading energy source. Next in importance, come coal and natural gas contributing almost 50% of the energy sources (Energy Information Administration 2007).

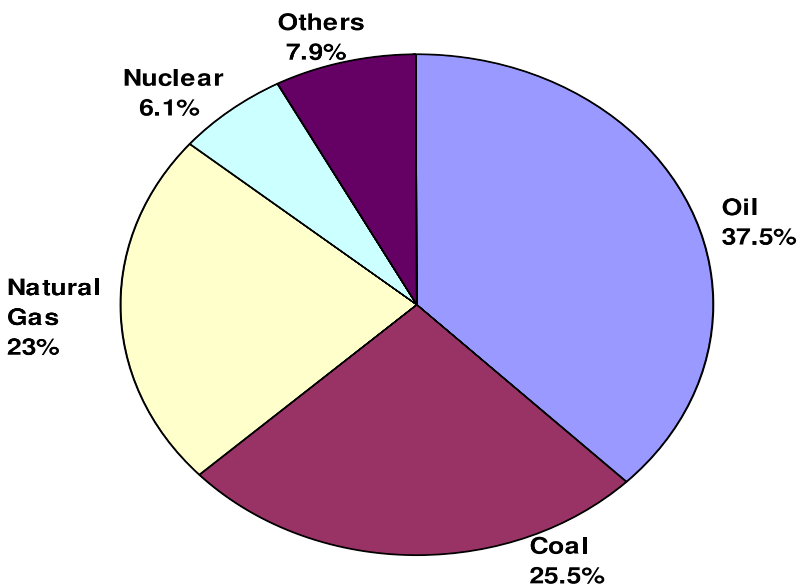

Figure 1.1 Primary sources of energy in the world in 2004. Total energy used was 446 Quadrillion Btu (Data were extracted from the Energy Information Administration 2007).

This thesis follows the style of Clean Technologies and Environmental Policy.

Natural gas is a clean source of energy and its popularity is expected to grow rapidly in the future because it presents many environmental advantages over oil and coal. Carbon dioxide (CO_2), a greenhouse gas related to global warming, is produced from oil and coal at rate approximately 1.4 to 1.75 times higher than that produced from natural gas. Also, nitrogen oxides (NO_x), greenhouse gas and a source of acid rain, are formed from burning fuel. NO_x produced from burning natural gas are approximately 20% less than those produced from burning oil and coal (Kidnay and Parrish 2006). In 2004, the worldwide consumption of natural gas was about 100 trillion cubic feet and is expected to grow 163 trillion cubic feet by the year 2030. Figure 1.2 shows the projected world total energy consumption by 2030 (Energy Information Administration 2007).

In 2006, the proven reserves of natural gas were reported to be 6,183 trillion cubic feet with the majority of these reserves being in the Middle East (2,566 trillion cubic feet) and Eurasia (2,017 trillion cubic feet). In fact, Russia, Iran, and Qatar combined account for about 58 percent of the world reserves. For instance, Qatar has a proven reserve (911 trillion cubic feet) of natural gas and the world's third largest supplier of natural gas with 15% of the global production (Energy Information Administration 2007). Table 1.1 provides a summary of the natural gas reserves in different countries.

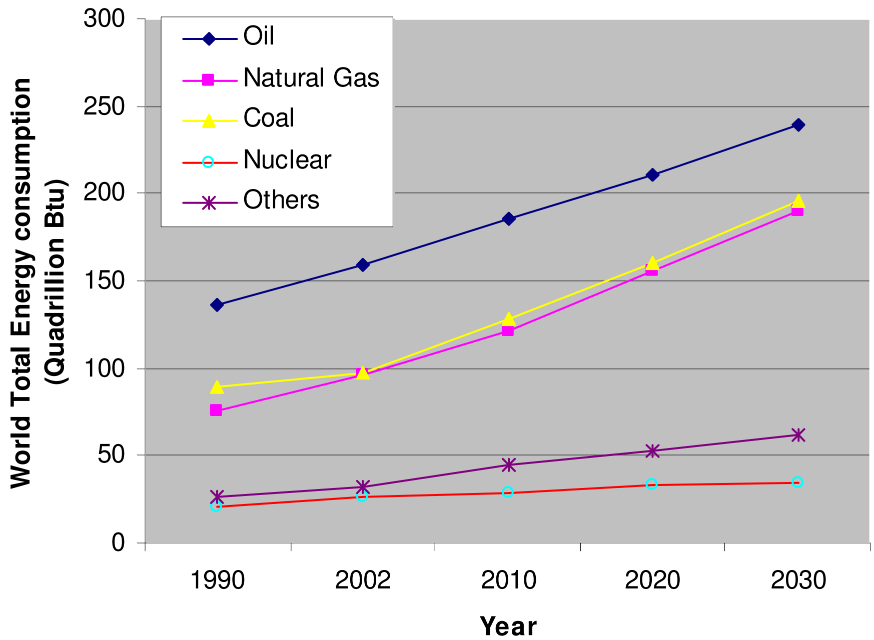

Figure 1.2 World energy consumption by fuel. (Data were extracted from Energy Information Administration 2007).

Table 1.1 World natural gas reserves by country as of January 1, 2007. (Data were extracted from Energy Information Administration 2007).

Country	Reserves (Trillion Cubic Feet)	Percent of World Total,%
Russia	**1680**	**27.2**
Iran	**974**	**15.8**
Qatar	**911**	**14.7**
Saudi Arabia	**240**	**3.9**
United Arab Emirates	**214**	**3.5**
United States	**204**	**3.3**
Nigeria	**182**	**2.9**
Algeria	**162**	**2.6**
Venezuela	**152**	**2.5**
Iraq	**112**	**1.8**
Others	**1352**	**21.8**
Total	**6183**	**100**

1.2 Overview of Liquefied Natural Gas (LNG) Market and Processing

The production of LNG has been practiced since the 1960's. The core concept in producing LNG is the condensation of natural gas. Other processing steps involve the elimination of undesirable impurities and separation of byproducts. Natural gas consists primarily of methane (CH_4), ethane (C_2H_6), propane (C_3H_8), small quantities of heavier hydrocarbons, nitrogen (N_2), oxygen (O_2), carbon dioxide (CO_2), water (H_2O), and sulfur compounds. Although the above-mentioned compounds may exist at low or high levels, CH_4 is the main constituent. A typical composition of natural gas is given in Table 1.2.

Table 1.2 Typical composition of natural gas mixture. A sample was taken from Qatar's North Field (Qatargas 2002).

Component	Mol%
H_2S	0.96
CO_2	2.45
N_2	3.97
CH_4	82.62
C_2H_6	4.84
C_3H_8	1.78
i-C_4H_{10}	0.39
n-C_4H_{10}	0.67
i-C_5H_{12}	0.29
n-C_5H_{12}	0.27
n-C_6H_{14}	0.34
Others	1.42
Total	100.00

Natural gas is liquefied to reduce its volume, increase its energy content (heating value) per unit volume, and facilitate energy transport in large quantities. Typically, LNG is stored and delivered at atmospheric pressure and -160℃ (-256℉). It is kept in specially designed storages and transported by special cryogenic sea vessels and road tankers. The density of LNG depends on its temperature, pressure, and composition. However, a typical density is about 500 kg/m^3. LNG is odorless, colorless, non-corrosive, and non-toxic. When vaporized in an air, LNG burns in the composition range of 5% to 15% (Center of Liquefied Natural Gas 2007). Neither LNG, nor its vapor, can explode in an unconfined environment.

As mentioned previously, methane is the principal element of natural gas. Accordingly, it is the primarily constituent of LNG. The final composition of LNG depends on the supplier, the feedstock, and the processing technology. The exact composition depends on the negotiation between the buyer and seller. Table 1.3 shows the composition range of the other constituents of LNG.

Table 1.3 Typical composition ranges of the LNG constituents. (Data were extracted from Kidnay and Parrish 2006).

Component	Composition Range (mol%)
Nitrogen	0.00-1.00
Methane	84.55-96.38
Ethane	2.00-11.41
Propane	0.35-3.21
Isobutane	0.00-0.70
n-Butane	0.00-1.30
Isopentane	0.00-0.02
n-pentane	0.00-0.04

International LNG trade is expanding rapidly. In the 1980's, only two grassroot LNG plants were built. In the 1990's, six grassroot LNG plants were built. To date, 15 large LNG plants had been built and are in operation. Furthermore, twenty-two LNG projects are reported to be planned with a total capacity of 110×10^6 metric ton per year (SRI Consulting 2003). In 2005, four countries (Indonesia, Malaysia, Qatar, and Algeria) accounted for approximately 60% of the world's exports of LNG. On the other hand, Japan and South Korea are the major LNG importers as they accounted for approximately 60% of world's LNG imports. Figures 1.3 and 1.4 show the 2005 worldwide LNG exporters and importers, respectively.

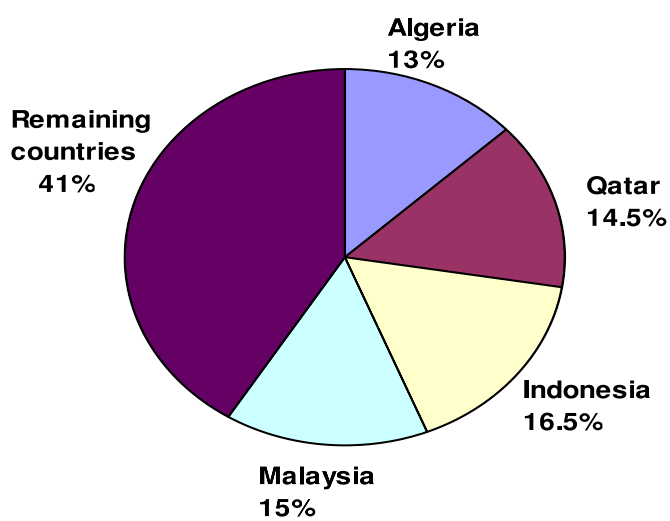

Figure 1.3 Worldwide LNG exports in 2005. Total exports were equivalent to 6.83 trillion cubic feet of natural gas (Data were extracted from Energy Information Administration 2007).

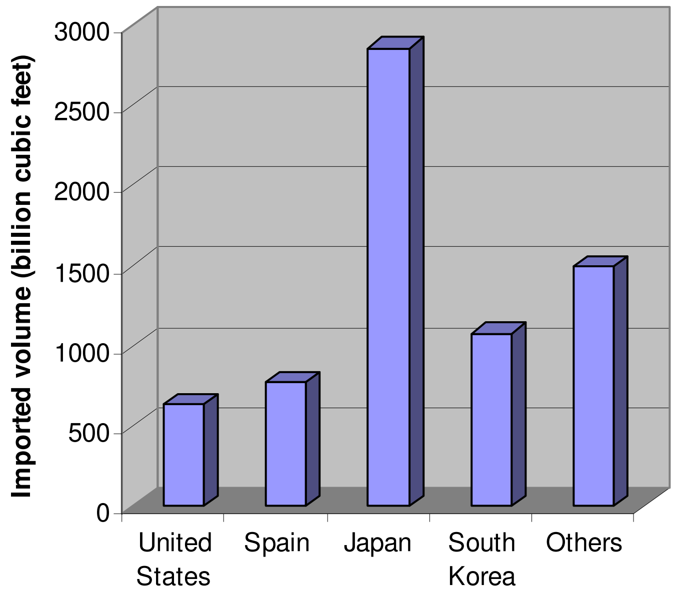

Figure 1.4 Worldwide LNG imports in 2005. Total exports were equivalent to 6.83 trillion cubic feet of natural gas (Data were extracted from Energy Information Administration 2007).

There are two common types of LNG plants:

- Peak shaving plants for seasonal production and storage
- Base load LNG plants for international trade with LNG shipped by the cryogenic sea vessels

The capacities of the peak shaving plants are smaller than base load LNG plants. Current and proposed base load LNG plants range in capacity from 1 to 15 $\times 10^6$ metric ton per year (SRI Consulting 2003).

Typically, an LNG supply chain involves four key steps: field processing of the produced raw natural gas, LNG production at the processing facility, shipping from the exporting country/location, and storage/regasification at the receiving terminals to be distributed as gas to end users. Figure 1.5 is a representation of these steps.

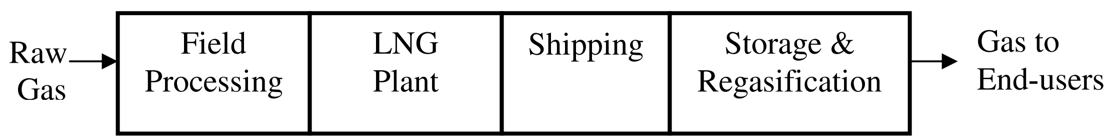

Figure 1.5 LNG Chain.

LNG is produced by a sequence of processing step that transform the natural gas from the gas phase to the liquid phase while getting rid of undesirable species. Figure 1.6 shows the block flow diagram of a typical LNG plant. It shows the key processing steps which start by gas treatment to remove acidic compounds followed by dehydration. Next, the desired hydrocarbons are recovered and nitrogen is rejected prior to final compression, liquefaction, and sales.

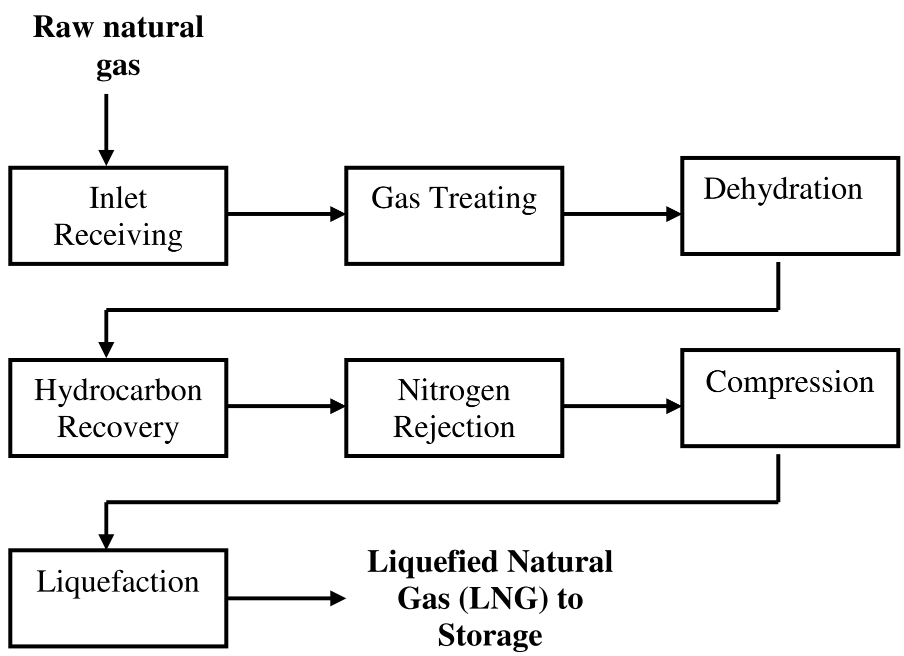

Figure 1.6 Block Flow Diagram of an LNG plant.

The inlet receiving units are designed, primarily for the initial gas-liquid separation. Additionally, condensed water, hydrocarbon liquids, and solids are removed. The other processing units are gas treating, dehydration, compression, and liquefaction. Gas treating involves reduction of the “acid gases” carbon dioxide (CO_2) and hydrogen sulfide (H_2S), to low levels such as 50 ppmv and 5 ppmv, respectively, to meet the standards and pipeline specifications. Dehydration involves removal of water and drying of the gas to avoid hydrate formation as well as corrosion. Liquefaction involves cooling the gas to an extremely cold temperature to convert it to liquid which basically represents the second part of the LNG chain.

1.3 Overview of the Thesis

The objective of this work is to assess, integrate, and optimize a typical LNG process. Section two provides a literature review on process integration in general and specifically as applied to the gas processing industry. Section three describes the problem statement. Section four gives the design approach and methodology. Section five presents the case study. Section six gives the results. Finally, the conclusions and recommendations are stated in section seven.

2. LITERATURE REVIEW

2.1 Process Integration

Process integration is a systematic approach to design, retrofit, and operate industrial facilities. It involves many activities such as, task identification, targeting, generation of alternatives, and analysis of selected alternatives. An important aim of process integration is to conserve process resources. These resources may be energy or material resources. Thus, process integration has been traditionally classified into energy integration and mass integration. Energy integration provides a systematic methodology to utilize the energy flow within the process. It covers all forms of energy including thermal, mechanical, electrical, and so on. On the other hand, mass integration provides a systematic methodology to utilize the mass flow, control, and optimize the species within the process. The fundamentals and applications of energy and mass integration have been reviewed in literature. Recently, a new category of process integration has been introduced. It is called property integration (El-Halwagi 2006). It provides a systematic design methodology which is based on properties and functionalities. Process integration is the combined and optimized version of the process synthesis and process analysis.

2.1.1 Process Synthesis

Synthesis means putting separate elements together. In process synthesis, we know process inputs and outputs and are required to design the process to reach the output for grassroot design or to revise the process for retrofitting design. For example, feed flowrate, composition, temperature, and pressure are given and the final product properties are also specified. We are required to design the various pieces of equipment, the different units and the whole plant in order to end up with that specified product. This activity is considered as process synthesis. Figure 2.1 shows the process synthesis block diagram.

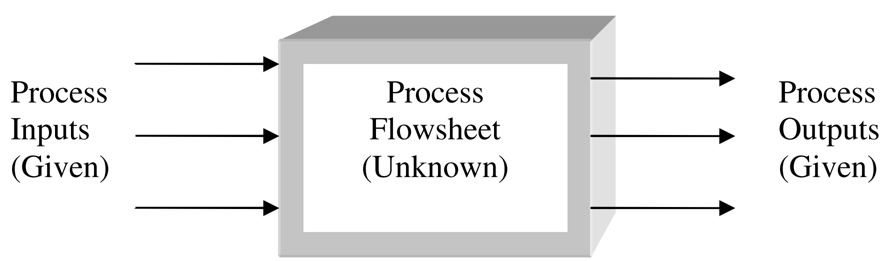

Figure 2.1 Process synthesis problem.

2.1.2 Process Analysis

Analysis means decomposition the whole to separate elements. So, it can be contrasted (and complemented) with process synthesis. In the process analysis we know process inputs and process flowsheet and we are asked to predict process outputs using different analysis techniques. These techniques include empirical correlations, mathematical models, and computer-aided process simulation tools. Figure 2.2 shows the process analysis block diagram.

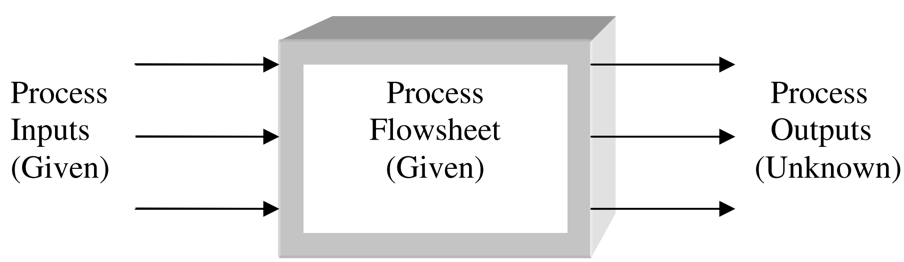

Figure 2.2 Process analysis problem.

2.1.2.1 Process simulation

Process simulation is one of tools used to predict the performance of the process. In this work, the process was simulated by using ASPEN Plus. It is a process simulation software package widely used in many applications. Given a process design and an appropriate selection of thermodynamic models, it uses mathematical models to predict the performance of the process. These models include material and energy balances, thermodynamic equilibrium, rate equations and so on. The predicted outputs might be stream properties, operating conditions, and equipment sizes. The computer-aided simulation has several advantages for example it:

- Allows the designer to quickly test the performance of proposed synthesized flowsheet
- Minimizes experimental efforts and scale up issues
- Helps obtaining optimum integrated design
- Explores process sensitivity by answering “what if” questions
- Sheds insights on process performance
- Tests the performance with various thermodynamics models

2.2 Energy Integration

As mentioned previously, there are two commodities utilized in the process: mass and energy. Both of them have the direct and indirect effects at process performance. These effects may be tangible such as savings in fixed capital and operating costs or intangible such as protecting the environment. At this section, both the heat integration and combined heat and power (CHP) integration will be discussed.

2.2.1 Heat Integration

Heat is one of the most important energy forms in the process. In a typical process, there are normally several hot streams that must be cooled and vise versa. The external cooling and heating utilities (e.g., cooling water, refrigerants, steam, and heating oil) should be provided to fulfill the process duties and needs. Using these external utilities in the process all the time is costly. Instead, integrating the process by transferring the heat from the hot streams to cold streams may lead to a significant cost reduction. This task is known as the synthesis of heat exchange networks (HENs) and as stream matching. The idea here is to utilize the available thermal energy before using external utilities. Thus, the objective of synthesizing HENs is the identification of minimum utility targets by using thermal pinch analysis.

Figure 2.3 shows the overall strategy for HEN synthesizing. There are three methods used for determining the minimum utilities required in HEN. The first method is the algebraic *temperature interval method* which was developed by (Linnhoff and Flower 1978). The second method is the *linear programming method*. The third one is the *graphical method using the pinch diagram.* Clearly, the determination of minimum utility requirements precedes the synthesis of HEN. The latter step or stream matching can be also done by one of the two following methods. The first one, which was introduced by (Linnhoff and Hindmarsh 1983), gives emphasis on positioning the heat exchangers by working out from the pinch temperatures. The second one, which was introduced by (Papoulias and Grossmann 1983), is an algorithmic strategy that utilizes a mixed integer linear program (MILP). Figure 2.4 shows the possible methods to synthesis HEN.

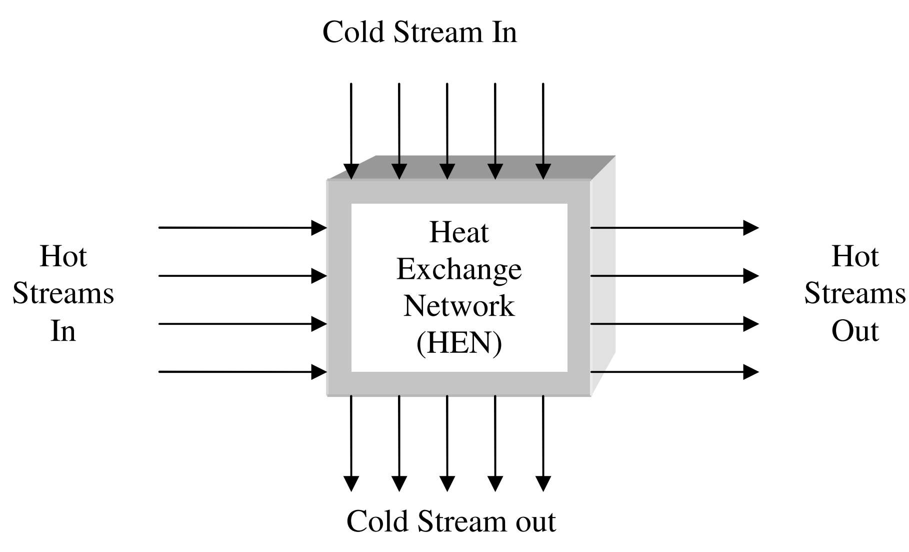

Figure 2.3 Synthesis of HENs.

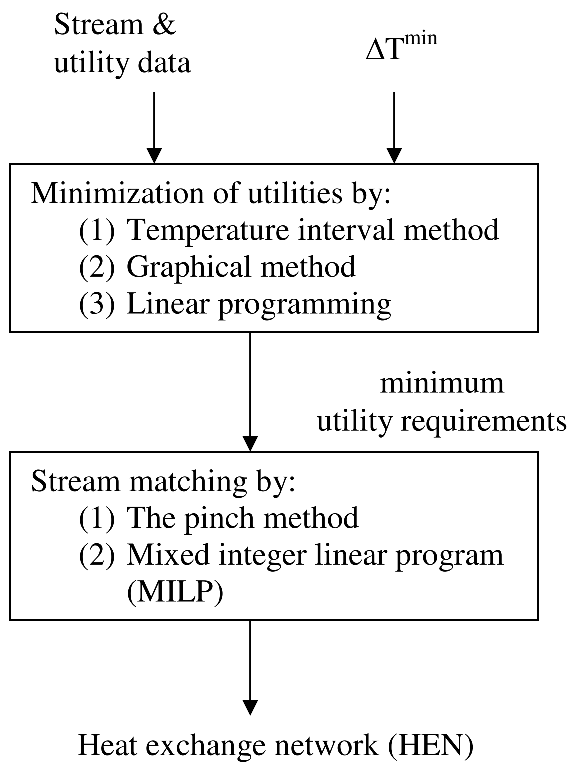

Figure 2.4 A possible two-stage method to the synthesis of HEN.

In this work, the *temperature interval method* will be used to find the minimum cooling and heating utility targets, and stream matching will be done by using the *mixed integer linear program (MILP)* solved in LINGO. However, for retrofitting the existing network of heat exchangers, a new method and optimization formulation will be developed.

2.2.1.1 Minimum utility target by the temperature interval method

In order to find the minimum utility target by the temperature interval method, some steps should be followed. (1) The temperature-interval diagram (TID) is constructed. TID is a useful tool for ensuring the thermodynamic feasibility of heat exchange. It has two temperature scales (hot and cold scales) and they can be determined by using equation (2.1), where T and t are the hot and cold scales respectively and ΔT^{min} is the minimum heat exchange driving force.

$$T = t + \Delta T^{min} \tag{2.1}$$

(2) Each stream is represented by a vertical arrow whose tail and head represent the supply and target temperature respectively.

(3) Horizontal lines are drawn at the heads and the tails of the arrows. These horizontal lines define a series of temperature intervals $z = 1,2,..,n$. It is thermodynamically valid to transfer heat from hot stream to cold stream, not only within the interval, but also, to any cold stream lies below that interval. (4) Table of exchangeable heat load (TEHL) is constructed. The exchangeable load of the u^{th} hot stream (losing sensible heat) which passes through the z^{th} interval is determined by:

$$HH_{u,z} = F_u\, C_{p,u}(T_{z-1} - T_z) \tag{2.2}$$

Where $HH_{u,z}$ is the hot load in interval z, F_u is the hot stream flowrate, $C_{p,u}$ is specific heat of hot stream u, T_{z-1} and T_z are the hot-scale temperature at the top and the bottom lines defining the z^{th} interval.

Similarly, the exchangeable load of v^{th} is determined by the following expression:

$$HC_{v,z} = f_v C_{p,v}(t_{z-1}-t_z) \tag{2.3}$$

where $HC_{v,z}$ is the cold load in interval z, f_v is the cold stream flowrate, $C_{p,v}$ is specific heat of cold stream v, $t_{z-1}$and t_z are the cold-scale temperature at the top and the bottom lines defining the z^{th} interval.

(5) Collective loads (capacities) of the hot (cold) are calculated by equations (2.4) and (2.5), respectively:

$$HH_z \text{ (total)} = \sum HH_{u,z} \tag{2.4}$$

$$HC_z \text{ (total)} = \sum HC_{v,z} \tag{2.5}$$

where $\sum$ represents the summing up of the individual load (capacity) of the hot (cold) stream. (6) Energy (heat) balance is performed around each interval. Figure 2.5 shows the heat balance around temperature interval z^{th}. r_o is zero, since no process streams exist above the first interval. (7) Finally, the most negative residual is added to the first temperature interval. This positive residual is the minimum heating utility requirement, $Q_{Heating}^{\min}$. A zero residual heat locates the thermal pinch location. The residual leaving the last interval is the minimum cooling utility requirement, $Q_{Cooling}^{\min}$.

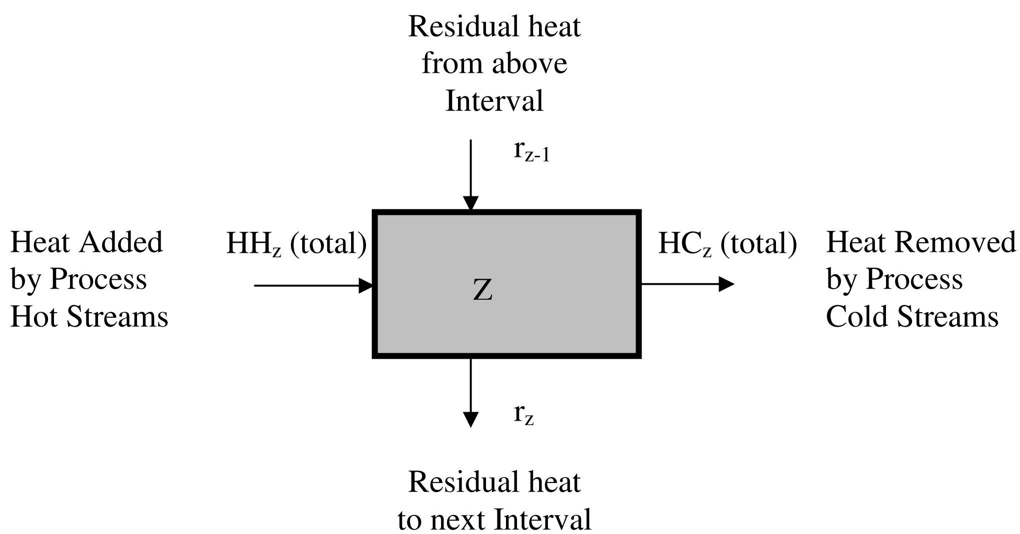

Figure 2.5 Heat balance around temperature interval.

2.2.1.2 Stream matching and network synthesis

Having determined the minimum utilities for heating and cooling and the values of all the individual and accumulative loads of hot and cold streams as well as the pinch location, we can now minimize the number of heat exchangers. It is common when a pinch point exists, the synthesis problem can be decomposed into two subnetworks, one above and one below the pinch temperature. The problem of minimizing the number of heat exchangers (Papoulias and Grossmann 1983) can be formulated as a mixed integer linear program (MILP):

minimize $$z = \sum_{m=1,2} \sum_{i \in R_m} \sum_{j \in S_m} E_{ijm}$$

Subject to the following constrains:

Energy balance for each hot stream around temperature intervals:

$$R_{ik} - R_{i,k-1} + \sum_{j \in S_{mk}} Q_{ijk} = Q_{ik}^{R} \qquad i \in R_{m,k} \ , k \in SN_m , m = 1,2$$

Energy balance for each lean stream around temperature intervals

$$\sum_{i \in R_{mK}} Q_{ijk} = Q_{jk}^{S} \qquad i \in S_{mk} \qquad k \in SN_m , m = 1,2$$

Matching of loads

$$\sum_{k \in SNm} Q_{ijk} - U_{ijm} E_{ijm} \leq 0 \qquad i \in R_m \quad j \in S_m , m = 1,2$$

Non negative residuals

$R_{ik} \geq 0 \qquad i \in R_{mk}, k \in SN_m, m = 1,2$

Non negative loads

$Q_{ijk} \geq 0 \qquad i \in R_{m,k}, j \in S_{mk}, k \in SN_m, m = 1,2$

Binary integer variables for matching streams

$Eijm = 0,1 \qquad i \in R_m, j \in S_m, m = 1,2$

The above program is an MILP that can be solved by using LINGO. It is worth noting that the solution of program will not be unique. It is possible to generate all integer solutions to it by adding constraints that exclude previously obtained solutions from further consideration. For example, any previous solution can be eliminated by requiring that the sum of $E_{i,j,m}$ that were nonzero in that solution be less than the minimum number of exchangers. For more details about the nomenclature of the MILP, the reader is referred to (Papoulias and Grossmann 1983) for heat exchanger units, and (El-Halwagi and Manousiouthakis 1990) for mass exchanger units.

2.2.2 Combined Heat and Power Integration

2.2.2.1 Cogeneration

Most of the process heating requirements are fulfilled by the steam. This steam is usually bought or produced at different levels i.e., high, medium, and low pressure. Each level has different cost per a unit of energy removal. From this regard, it is better to produce the steam at high pressures and pass it through a turbine to end up with both the required medium or low pressures and the produced work or the shaft work. This process is referred to as cogeneration or combined heat and power. The generated power can be obtained from Mollier diagram as shown in Figure 2.6. For a given inlet pressure and temperature, and a given outlet pressure of the turbine, the isentropic enthalpy change in the turbine can be determined as:

$$\Delta H^{isentropic} = H^{in} - H_{is}^{out}$$

where $\Delta H^{isentropic}$ is the specific isentropic enthalpy change in the turbine, H^{in} is the specific enthalpy of the steam at the inlet temperature and pressure of the turbine and H_{is}^{out} is the specific isentropic enthalpy at the outlet pressure of the turbine. An isentropic efficiency term is defined as follows:

$$\eta_{is} = \frac{\Delta H^{real}}{\Delta H^{isentropic}}$$

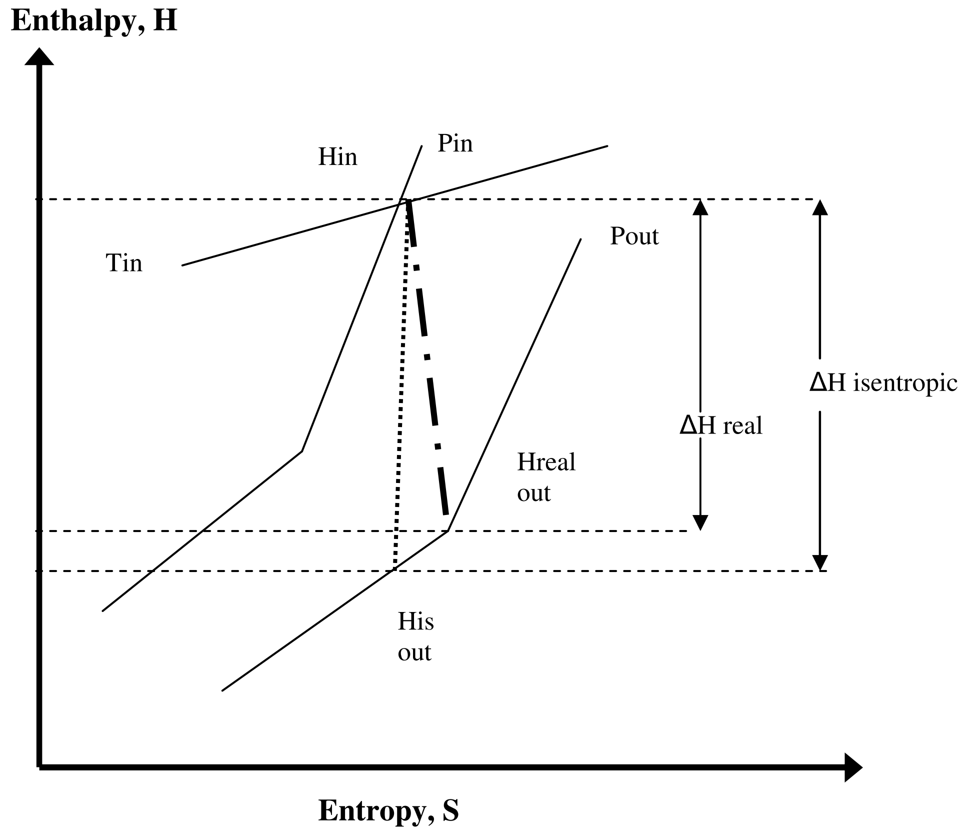

Figure 2.6 Turbine power represented on the Mollier diagram.

where η_{is} is the isentropic efficiency and ΔH^{real} is the actual specific enthalpy difference across the turbine.

Also, the power produced by the turbine, W, is given by:

$$W = F\eta_{is}(H^{in} - H_{is}^{out})$$

$$F \cong \frac{Q}{\Delta H_{lv}}$$

where m is the steam flowrate, Q is heat duty and ΔH_{lv} is the enthalpy difference between saturated liquid and vapor. The last equation is beneficial to target cogeneration potential for a process without detailed calculations.

2.2.2.2 Turbo-expander in natural gas plants

Turbo-expanders have proven high level of reliability in many power recovery applications. Theses include natural gas processing, oil field cogeneration, and recovery of energy from waste heat. Turbo-expanders have been used in oil and gas industry for many years as a cryogenic tool. They are primarily used in natural gas processing to produce low temperature refrigeration needed for the hydrocarbons recovery. Turbo-expanders are also used to lower the pressure of the gas stream. Accordingly, the gas is cooled significantly while the turbo-expander produces work. The amount of produced power depends on the gas flow rate and the expansion pressure ratio (head). Typically, the ratio is 4 to 1. The produced power is usually used in a direct driven compressor.

Turbo-expanders are composed of the turbine and its associated units such as, compressor, separator, and Joule-Thomson value. Turbo-expanders are used in natural gas processing plant particularly, in fractionation unit. The natural gas is cooled to extremely low temperature. After that the cold liquid and vapor are separated in a low temperature separator (LTS). The liquid stream is flashed across a Joule-Thomson (JT) valve for pressure reduction and additional cooling and fed to the distillation column to replace the reflux. The vapor stream of the LTS is fed to the turbine where temperature is reduced and the work is produced. Further separation of liquid produced by the expander from vapor takes place in vapor liquid separator (VLS). Also, the produced liquid is fed to top of the column to serve as reflux. The combined vapor stream from VLS and the column is fed to liquefaction unit for further cooling requirement. The energy of bottom product is utilized by cooling the incoming natural gas stream. Figure 2.7 shows a simplified flowsheet of turbo-expander process.

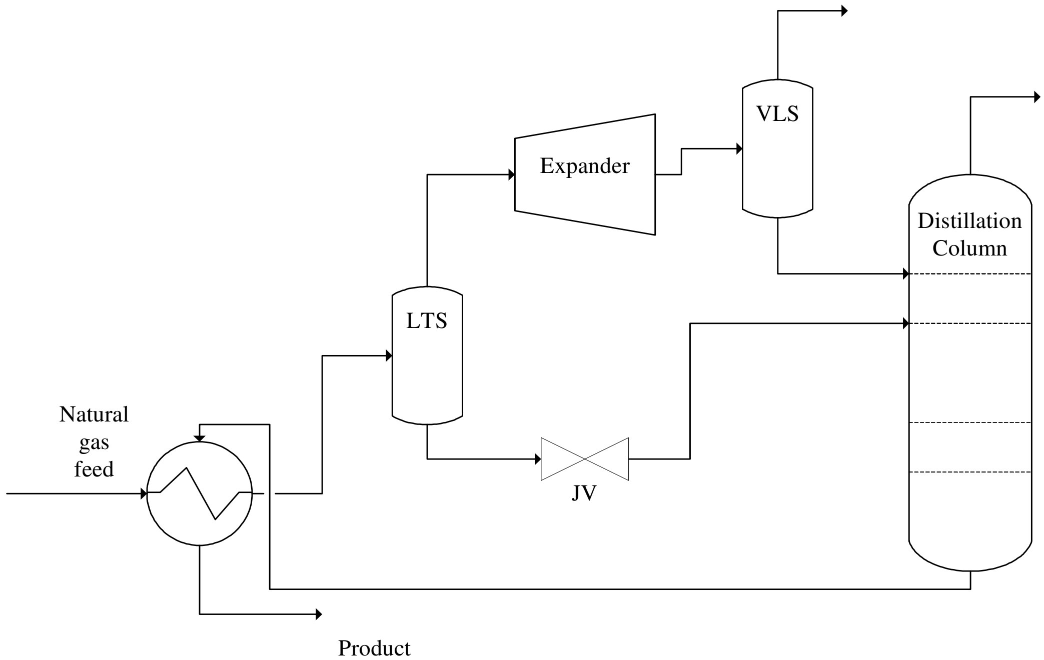

Figure 2.7 Simplified flowsheet of turbo-expander process.

2.3 Mass Integration

By the end of 1980’s, another process commodity was considered. (El-Halwagi and Manousiouthakis 1989) developed and designed mass exchange networks (MEN). Mass integration is a systematic approach that deals with the material flow within the system. It involves many activities, such as targeting, recycling, mixing, and synthesizing of mass exchange networks (MEN). For more details the reader is referred to (El-Halwagi 2006).

2.4 Earlier Work in Natural Gas Integration

This subsection provides a selective review of earlier work in the area of process integration activities for the optimization of LNG facilities.

2.4.1 Integrated Liquids Recovery

(Hudson et al. 2004) studied the recovery of liquids (hydrocarbons) recovery through fractionation. They concluded that integration of this unit will improve the overall LNG production efficiency. Their main concern was splitting the low level refrigeration to supply a portion of the process cooling. Also, they considered the work expansion machines (i.e., turbo-expanders) to generate part of the refrigeration. Finally, they concluded that this integrated recovery process can be applied to any type of gas liquefaction cycle and can be adapted for any changes in the feed gas composition.

2.4.2 Flexibility Analysis in Designing a Heat-Exchanger Network (HEN)

(Konukman and Akman 2005) examined the synthesis of the HEN in a natural gas turboexpander plant for ethane recovery. Their key contribution was the includion of operability and flexibility index computations. Their emphasis was given to the flexibility of the process in maintaining a certain level of ethane recovery under variations of natural gas feed stream temperature and pressure.

2.4.3 Optimum Waste Interception

(Hamad et al. 2007) studied the acid gas removal unit which represents the gas treating unit in this study of an LNG facility. Their key focus was the reconciliation of environmental and energy objectives. They concluded that considering energy integration opportunities in the process will lead to an optimum waste interception network. Also, they highlighted the importance of relating the objectives of mass and energy integration.

2.5 LNG Process Description

2.5.1 Inlet Receiving Unit

The feed gas usually comes from high pressure sub-sea pipeline. The gas from offshore enters the inlet receiving facilities at the slug catcher and is stabilized at the condensate stabilization units.

2.5.1.1 The slug catcher

The primary gas/liquid separation takes place in the slug catcher. It is a multiple pipe finger type and is sized to retain liquid slugs arriving from transient and pigging operation of the long sea pipeline. The main body of the slug catcher consists of gas/liquid separation fingers cross connected to each other. The gas/liquid separation fingers are sloped down to the liquid outlet header.

2.5.1.2 The condensate stabilization unit

The condensate from the slug catcher flows to the flash drum which provides three phase separation: gas, condensate and water. Condensate from the bottom of the flash drum is fed to the top tray of condensate stripper. The gas leaving the stripper is fed to the low pressure stage of gas compressor. The last tray temperature of the stripper is controlled by the fuel gas firing to the furnace. The stripper reboiler is fuel-fired to provide bottoms temperature that meets Reid Vapor Pressure (RVP) specification. The gas from the flash drum and condensate stripper is compressed in a two-stage centrifugal compressor and is subsequently combined with the gas from the slug catcher to provide the feed gas to gas treating unit.

2.5.2 Gas Treating Unit

The gas treating unit involves the removal of acid gases to low levels. The removal task mainly targets H_2S and CO_2 from the gas stream. The removal task may be carried out chemically, physically, or through a combination of them. The decision of which type of process should be selected depends on several issues. Some of them are summarized here from (Gas Processors Suppliers Association 1994):

- Type and concentration of impurities in the sour gas
- Specifications for residue (treated) gas, the acid gas, and liquid products.
- Volume of the processed gas
- Capital and operating costs

2.5.2.1 Chemical reaction process

It involves the removal of acid gases in two steps. First, the gas is absorbed by the solvent. Then, it reacts chemically with the material in the solvent. Amines have been commonly used for this purpose. Amines are formed by replacing one or more hydrogen atoms from ammonia (NH_3) with a hydrocarbon group. These amines are used in water solutions at different concentrations. There are many aqueous alkanolamine processes. For example, monoethanolamine (MEA) is commonly used in acid treating. Others might be diglycolamine (DGA), diethanolamine (DEA), methyldiethanolamine (MDEA) and so on. Each one of them has some advantages and disadvantages in using over other amines and has its own operating parameters. The operating parameters include the required concentration, rich amine acid gas loading and etc.

In the acid gas unit, the gas is treated by the solvent (lean amine) in a contactor (absorber) at high pressure. However, the rich amine (amine with dissolved gases) is regenerated in a low-pressure stripper.

2.5.3 Dehydration Unit

The dehydration process involves the removal of water. It prevents the formation of hydrates and corrosion. Also, it helps in meeting specifications on water content. There are several methods used to dehydrate natural gas. The most commonly used techniques include absorption using liquid desiccants, and adsorption using solid desiccants.

2.5.3.1 Absorption process

In the absorption process the water is contacted with a liquid desiccant. Mostly, the glycols are the most used absorbents. These absorbents could be ethylene glycol (EG), diethylene glycol (DEG), and triethylene glycol (TEG).

In the glycol absorption unit, the gas flows upward the contactor and is absorbed by the lean glycol solution. The rich glycol absorbs the water and leaves at the bottom of the column. Then the rich glycol is heated to regenerate the lean glycol. The operating conditions of the glycol unit are controlled according to the type of glycol, and required water removal. For more information about the detailed design of glycol unit and typical operation conditions of each type of glycol, and the adsorption processes the reader is referred to the (Gas Processors Suppliers Association 1994).

2.5.4 Fractionation Unit

The fractionation unit involves the separation of heavy hydrocarbons into individual products. This removal is considered as a part of feed conditioning. There are several reasons for this step such as (1) controlling of LNG heating value (2) selling the recovered hydrocarbons as by-products (3) protecting the equipment from plugging due to freezing at low temperatures. This step is done by using the distillation columns (fractionators). There are types of fractionators. According to the products or achieved target, the fractionators are named. A de-methanizer refers to a distillation column where the methane is separated from the rest of hydrocarbons. Similarly, the de-ethanizer, de-propanizer, and de-butanizer are named for the removal of ethane, propane, and butane, respectively. The fractionators differ not only in their goals, but also in their internals. These internals might be trays or packing material. Also, another classification is attributed to the type of the condenser: total or partial condenser. In the case of the total condenser, all the vapor leaving the top of the column is condensed to liquid and return back to the column as a reflux. In the case of the partial condenser, a portion of the vapor is condensed as a reflux. In some cases, there is on condenser and the best example of that is the de-methanizer in the cryogenic plants where the reflux is replaced by a liquid feed stream.

2.5.5 Liquefaction Unit

The liquefaction unit involves the condensing of the natural gas. Gas can be condensed by (a) pressure effects (such as the ones using the Joule-Thomson cycles (b) compressing and expanding it using an engine doing external work like a turbo-expander (c) refrigerating it. A common configuration is for the gas to be cooled through the cold box by using a mixed refrigerant. For more details about the cycles the reader is referred to (Gas Processors Suppliers Association 1994).

3. PROBLEM STATEMENT

Consider a base-case process for the production of LNG with given input data, product specifications (e.g., product purity, heating value, total inert, sulfur content, etc.), and process constraints. It is desired to:

- Simulate the performance of the process to determine the characteristics and interactions of the key pieces of equipment and streams and to evaluate the process requirements of utilities
- Optimize design and operating conditions of the separation system
- Conduct process integration improvements to reduce energy consumption of the process. For instance, identify benchmarks (targets) for the minimum usage of external heating and cooling utilities and retrofit the existing network of heat exchangers in the process to attain the desired targets
- Evaluate opportunities for improving the performance of the refrigeration system of the process

- Analyze and optimize the combined heat and power issues of the process and examine the prospects of process cogeneration

Figure 3.1 is a schematic representation of the stated problem.

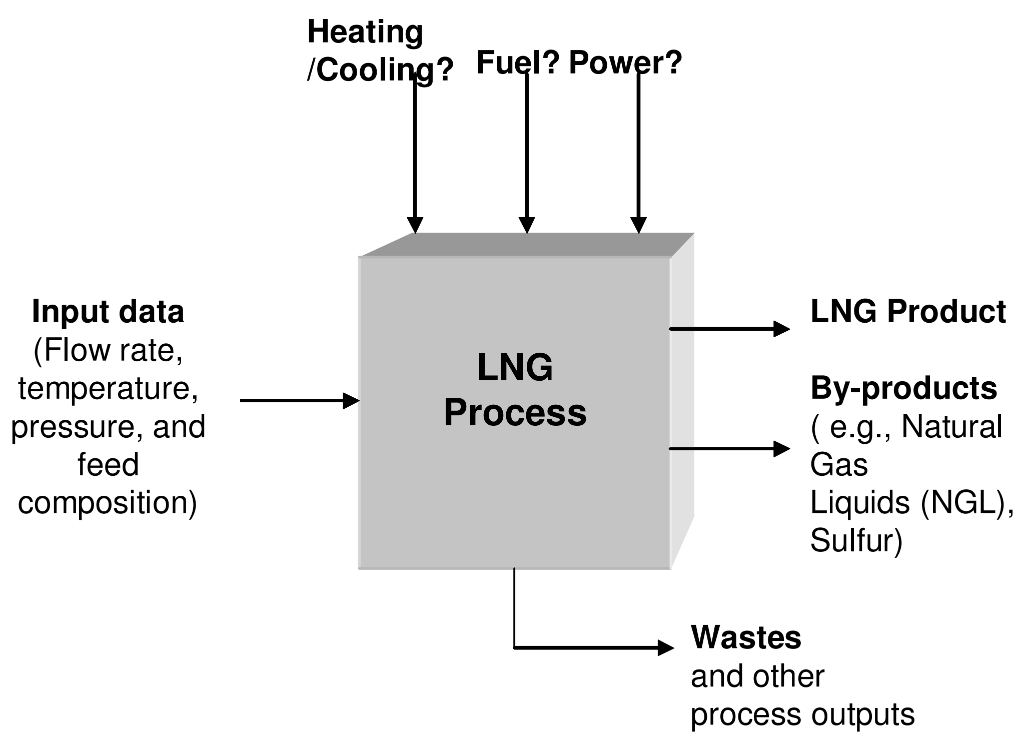

Figure 3.1 A schematic representation of the stated problem.

4. METHODOLOGY AND DESIGN APPROACH

4.1 Overview of the Design Approach

The design approach is aimed at analyzing and improving the performance of the LNG process. Given the significant energy usage in LNG production, special emphasis is given to enhancing the energy metrics of the process. In this regard, the following four activities are undertaken to reduce energy consumption and enhance the process performance:

- Optimization of the design and operating variables of the process
- Heat integration and retrofitting of heat exchangers
- Process cogeneration for optimizing combined heat and power
- Assessing the usage of turbo-expanders for efficient refrigeration

Figure 4.1 is an illustration of these four activities.

The following is a description of the proposed methodology and design approach. It is a hierarchical procedure consisting of sequential steps. The first step is the construction of a typical process flowsheet which is based on widely-used LNG technology. The second step is the simulation of the synthesized flowsheet. A computer-aided simulation package (ASPEN Plus) is used to determine the steady-state characteristics of the key units and streams for a base-case process. Before running the simulation, an appropriate thermodynamic property method (e.g., Peng Robinson) is selected. Furthermore, studies on design specification and sensitivity analysis are performed at this step.

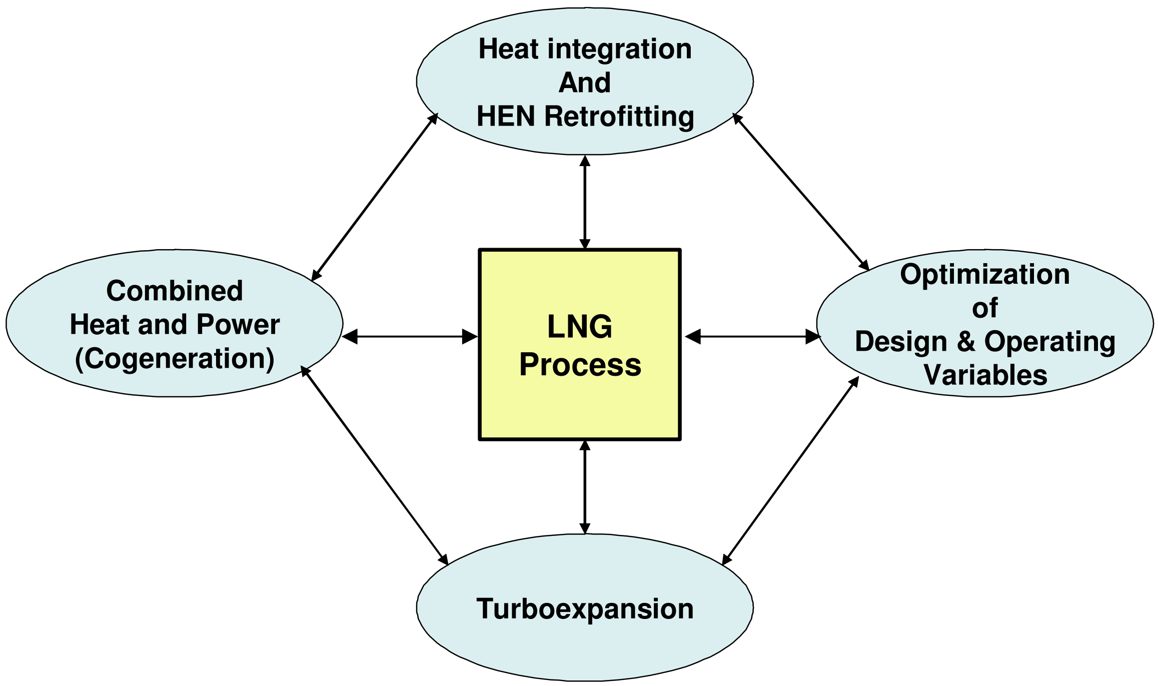

Figure 4.1 Proposed activities for process and energy improvement.

The next step is to modify the process configuration to reduce heating and cooling utilities. Focus is given to the fractionation train and its potential precooling. This is an important activity aimed at the optimization of the process, particularly the fractionation train since it is the largest consumer of energy. In this regard, there is a need to reconcile the heating and cooling utilities, the types and quantities of the various utilities, and the sizes of the columns. The third step includes the implementation of a thermal pinch analysis by which the minimum heating and cooling utilities for the process are obtained. In order to identify an actual network attaining these targets, the task of synthesizing a network of heat exchangers has been formulated as a mixed integer linear program (MILP) and solved using LINGO (an optimization software package). The fourth step includes the retrofitting of the heat exchangers through a novel approach which is described in the following section. The last step includes the application of cogeneration activity. A schematic representation of these steps is shown in Figure 4.2.

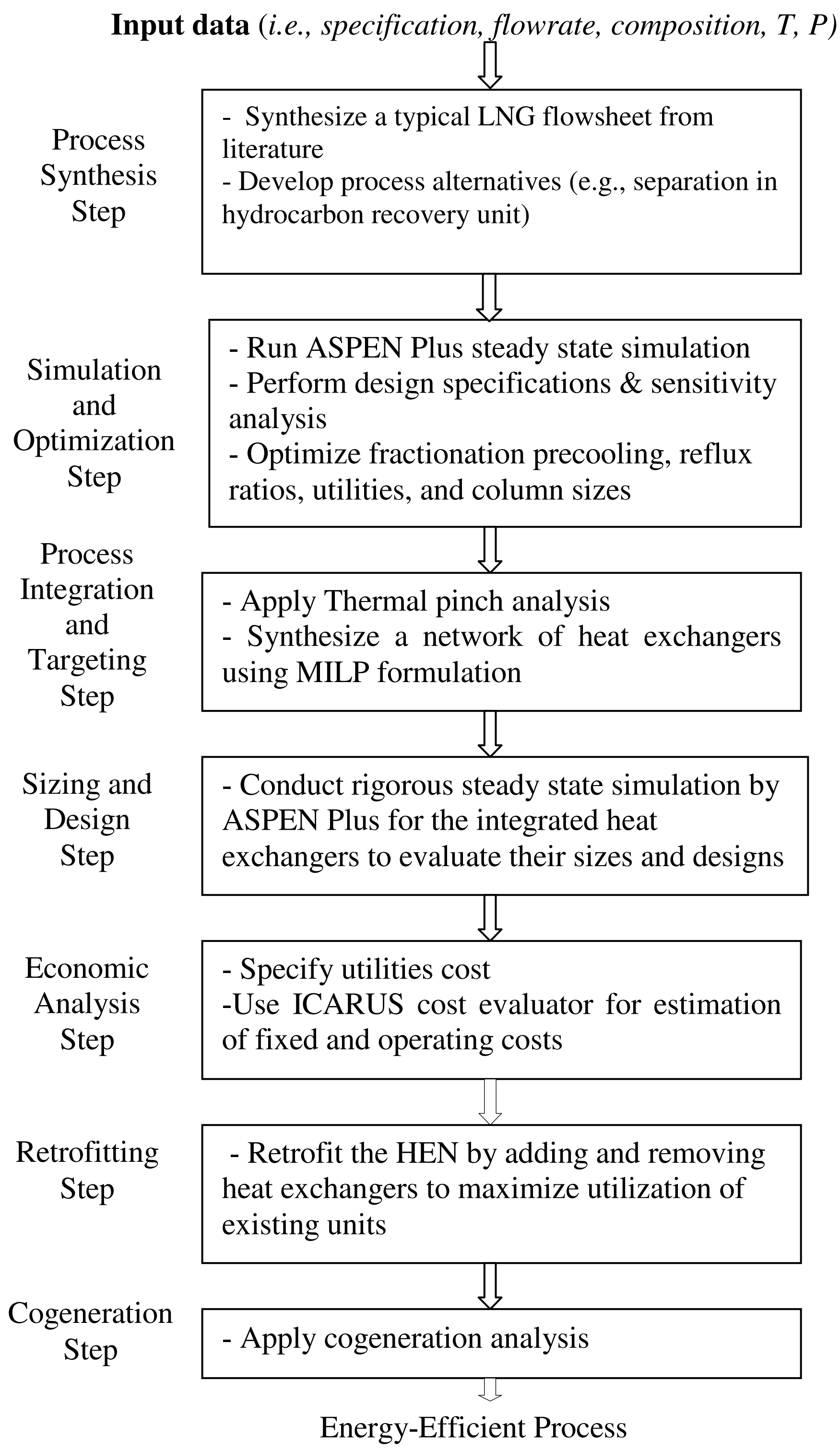

Figure 4.2 Overall methodology and design approach for heat integration and cogeneration strategies.

4.2 Novel Approach and Mathematical Formulation for HEN Retrofitting

The retrofitting of the existing HEN is a challenging task. The majority of the literature techniques are on the synthesis of a grassroot HEN. The retrofitting literature for HEN is mostly geared towards conceptual design with order-of-magnitude estimates. Simplifying assumptions on heat transfer coefficients and unit sizing are often used. These assumptions may lead to inaccurate estimates of the required heat-transfer areas and may not lead to feasible implementations. Our objective is to include rigorous simulation and sizing in the retrofitting analysis. The difficulty is that there are three platforms to be integrated: (1) process integration for minimizing heating and cooling utilities and for generating HEN matches (without linkage to the existing heat exchanger units). (2) Process simulation for the rigorous sizing of the integrated heat exchangers. (3) Process optimization for assigning the integrated exchangers to the existing units and for identifying new units to be added. The guiding objective is to minimize energy consumption while maximizing the utilization of the existing heat exchangers in implementing the identified heat integration strategy so as to decrease the overall cost of the project. Therefore, the thermal pinch analysis is first used to determine minimum heating and cooling utilities and an MILP optimization formulation is used to develop a network of heat exchangers with minimum number of units that satisfy the minimum heating and cooling utilities. Integer cuts are used to generate the various network alternatives. The generated matches for heat exchangers are simulated using ASPEN Plus to obtain rigorous sizing. Now, we have the idea of grassroot HEN along with its simulation and sizing of the heat exchangers. The question is how to maximize the utilization of the existing heat exchangers for retrofitting. Towards this end, the following new approach and optimization formulation are developed:

Now, we have the actual heat transfer area for both the existing and integrated heat exchangers. In order to utilize the existing units, the retrofitting step will be carried out. The retrofitting step can be done by inspection, but the optimal solution is not guaranteed particularly for a large number of heat exchangers. Therefore, systematic approach should be developed to aid in the retrofitting of the HEN. While useful research has been published in the area of retrofitting HENs, they have one or more of the following limitations:

- Oversimplified models for sizing the heat exchangers (e.g., fixed value of overall heat transfer coefficient regardless of the hot and cold streams or the exchanger geometry, pre-determined heat-exchanger types, etc.)
- Subjective rules for matching integrated heat exchangers with existing units
- Complicated mathematical formulations that may not be globally solvable

Our objective here is to develop a novel HEN retrofitting procedure that enjoys the following features:

- Systematically identifies matches between the integrated heat exchangers and the existing units
- Considers the actual properties of the hot and cold streams
- Allows rigorous simulation and sizing of heat exchangers

Towards this objective, the following procedure is proposed:

1. A rigorous simulation is carried out for the current process including all heat exchangers. This will result in identifying the heat loads, conditions, and sizes of the heat exchangers and will enable data extraction to set up the HEN synthesis problem for heat integration.
2. The pinch analysis is conducted to minimize heating and cooling utilities.
3. A grand composite curve is developed to screen and select the utilities and their loads. It is worth noting that steps 2 and 3 may be substituted by a linear program that determines the minimum heating and cooling utilities as well as the optimal load of each utility.
4. An MILP formulation is developed to determine the minimum number of exchangers satisfying the minimum heating and cooling utilities. The result is the network configuration along with heat-exchange loads, flows, and temperatures of hot and cold streams.
5. A rigorous simulation is carried out to model and size the various exchangers in the identified integrated HEN.
6. The integrated heat exchangers are matched with existing exchangers and the number and size of the new heat exchangers are determined. In order to carry out this step systematically, a new mathematical formulation is developed. It is described in the following section.

Figure 4.3 shows the proposed procedure for retrofitting HEN.

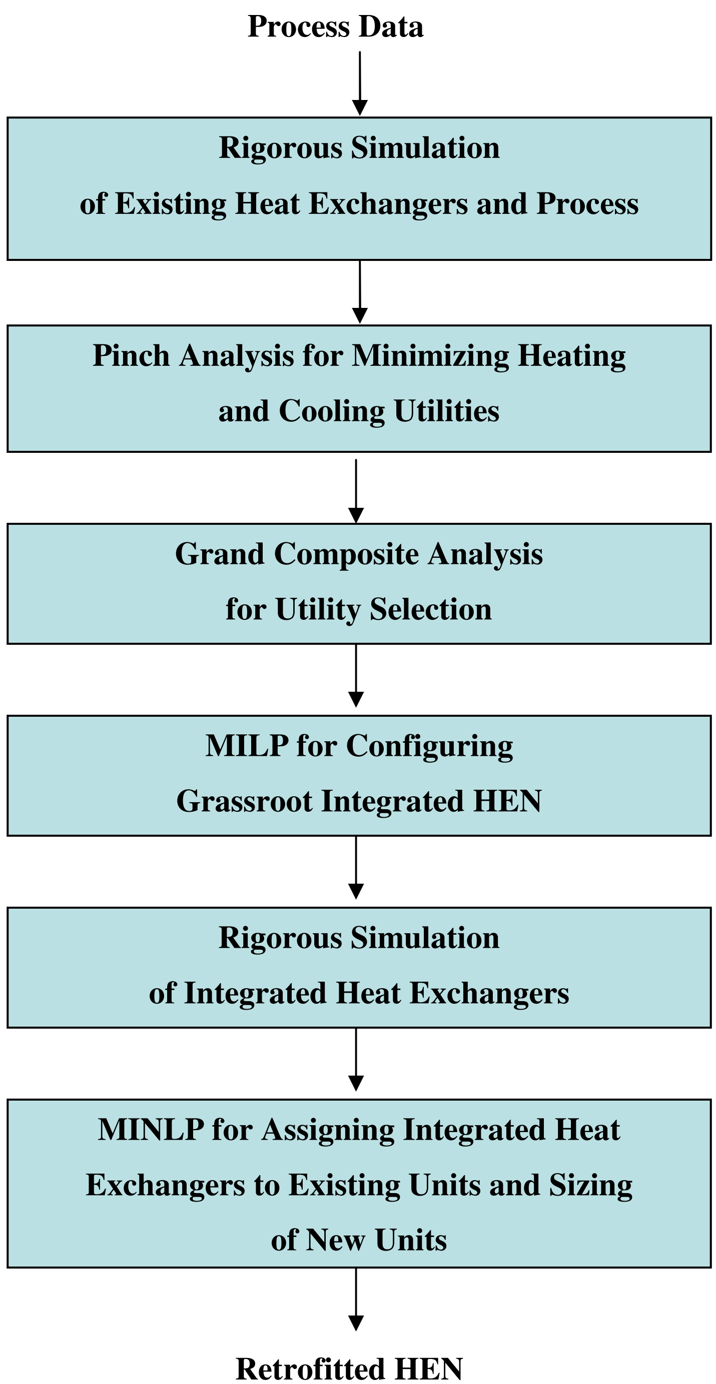

Figure 4.3 Proposed procedure for retrofitting HEN.

The following is a description of the new approach for retrofitting the HEN. First, a structural representation as shown in Figure 4.4 is developed to embed potential configurations of interest. Let us define the following sets:

- Existing_ Exchangers = {j|j=1, 2, …, $N_{existing}$} is the set of existing heat exchangers already in the process. The area of the jth heat exchanger, $A_j^{existing}$, is known (as determined by rigorous simulation of the process before integration).
- Integrated_ Exchangers = {i|i=1, 2, …, N_{integ}} is the set of integrated heat exchangers identified from the grassroot analysis of heat integration. The area of the ith heat exchanger, A_i^{integ} , is known (as determined by rigorous simulation).
- New_ Exchangers = {i|i=1, 2, …, N_{integ}} is the set of new heat exchangers to be added to supplement the use of existing heat exchangers. The area of the ith heat exchanger, A_i^{new} , is unknown and will be determined so as to minimize the cost of the retrofitted HEN.

The structural representation assigns each integrated heat exchanger to all existing heat exchangers to describe potential matches between the ith integrated heat exchanger and any of the existing heat exchangers. Such an assignment is modeled through 0/1 binary integer variables that will determine whether or not a match exists. These binary variables are defined as follows:

$E_{ij} = 1$ if there is a match between integrated exchanger i and existing exchanger j

$E_{ij} = 0$ if there is no match between integrated exchanger i and existing exchanger j

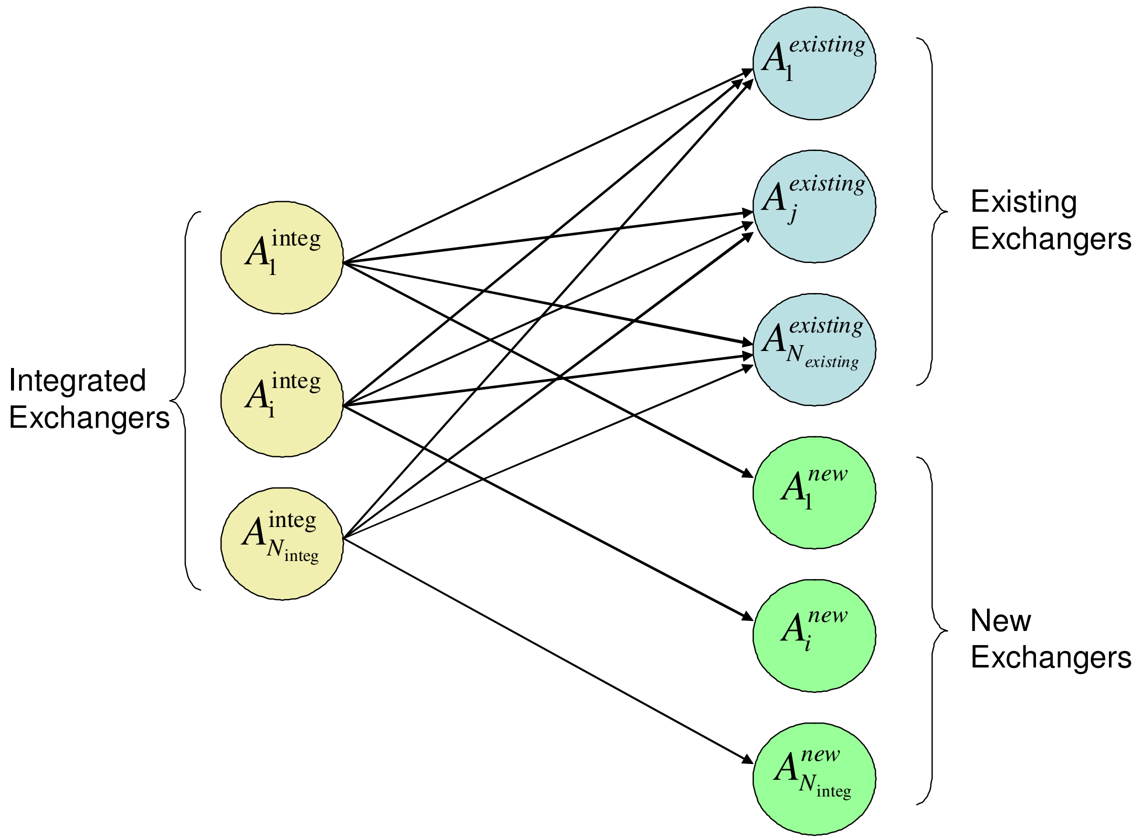

Figure 4.4 Structural representation of new retrofitting approach.

Additionally, each integrated exchanger i is assigned to a single new heat exchanger to be added if existing heat exchangers are not sufficient for the integrated exchanger.

Such an assignment is modeled through 0/1 binary integer variables that will determine whether or not a new exchanger is needed. These binary variables are defined as follows:

$E_i = 1$ if there is a need for a new heat exchanger for the ith integrated heat exchanger

$E_i = 0$ if there is no need for a new heat exchanger for the ith integrated heat exchanger

Consider the fixed cost of a heat exchanger to be given by:

$$FC_i = a_i + b_i * A_i^{new^x}$$

where a_i is a constant reflecting the fixed charges associated with the site-preparation work involved in the installation of a new heat exchanger exchanger and the constants b_i and x (usually x ~0.6) are used to evaluate the fixed cost of the exchanger.

Therefore, the following new formulation is proposed:

The objective function is:

$$\text{Minimize } \sum_i (a_i + b_i * A_i^{new^x}) * E_i$$

If there are no fixed charges for each new exchanger, then the objective function is simplified to:

$$\text{Minimize } \sum_i \ b_i * A_i^{new^x}$$

Furthermore, if the objective is to minimize the heat transfer area of the new exchangers (instead of the fixed cost of the new exchangers), then the objective function becomes the following linear expression:

Minimize $\sum_{i} A_i^{new}$

Subject to the following constraints:

Matching of Areas:

$$A_i^{new} + \sum_j E_{ij} * A_j^{existing} >= A_i^{\text{integ}} \qquad \forall i$$

This constraint stipulates that the area of the integrated heat exchanger may be distributed over existing heat exchangers and the remaining required area must be assigned to a new exchanger to be installed. The inequality sign indicates that at least this area is needed for the integrated exchanger.

Assignment of Existing Exchangers:

$$\sum_i E_{ij} <= 1 \qquad \forall j$$

While each integrated exchanger may be retrofitted using multiple existing exchangers, each existing exchanger may be matched to at most one integrated exchanger. This is attributed to the practical difficulty of retrofitting an existing exchanger to replace multiple exchangers.

Assignment of New Exchangers:

$$A^L * E_i \le A_i^{new} \le A^u * E_i \qquad \forall i$$

Where A^L is a given number representing a lower bound on the area of the exchanger, below which it is not practical to install a new exchanger and A^u is a given number representing an upper bound on the area of the new exchanger. This constraint also serves to assign the value of binary variable E_i to one or zero depending on whether or not there is a need for the new area of the heat exchanger. The LINGO code formulation for the case study and its solution can be found in appendix C.

Next, the economic factors of the retrofitted HEN are evaluated. Fixed capital and operating costs are estimated. The fixed capital investment for the newly added equipment is obtained by using ICARUS (cost evaluation software package).

5. CASE STUDY

5.1 Products Specifications and Design Basis

5.1.1 Feed Conditions

The following are feed gas conditions and Table 5.1 shows the base-case feed composition in mol%.

- Flow rate:1,500 MMSCFD
- Temperature: 20 ℃ (68 ℉)
- Pressure: 80 bara (1176 psia)

Table 5.1 Feed composition (dry basis) mol%.

Component	Mol%
H_2S	0.96
CO_2	2.45
N_2	3.97
CH_4	82.62
C_2H_6	4.84
C_3H_8	1.78
$i\text{-}C_4H_{10}$	0.39
$n\text{-}C_4H_{10}$	0.67
$i\text{-}C_5H_{12}$	0.29
$n\text{-}C_5H_{12}$	0.27
$n\text{-}C_6H_{14}$	0.34
Others	1.42
Total	100.00

5.1.2 Sweetened and Dehydrated Gas

The following are conditions and specifications for the sweetened and dehydrated gas:

- Temperature: 45 ℃
- Pressure: 67 bara
- CO_2 content : 50 parts per million by volume (ppmv)
- H_2S content : 5 ppmv
- Water content : 0.1 ppmw

5.1.3 Liquids Products

Table 5.2 shows the specification of the fractionation products (ethane, propane, and butane from the de-ethanizer, the de-propanizer, and the de-butanizer, respectively).

Table 5.2 Liquid products specifications.

	Ethane	Propane	Butane
Methane	**2 mol%**	**1 ppmw**	**0**
Ethane	**96 mol%**	**2 mol %**	**1 ppmw**
Propane	**2 mol%**	**95 mol%**	**2 mol%**
Butanes	**0**	**3 mol %**	**97 mol%**
Pentanes	**0**	**15 ppmw**	**1 mol%**
H_2S	**53 ppmv**	**0**	**0**

5.1.4 LNG Product

The following are the desired specifications for the LNG product:

- Temperature: -160 °C (-256 ° F)
- Pressure: 1 bara
- Methane : Minimum 85 mol%
- Propane : Maximum 3.5 mol%
- CO_2 content : Maximum 50 parts per million by volume (ppmv)
- H_2S content : Maximum 3 ppmv

5.2 Utilities Specifications

Several utilities are used for heating and cooling. Their conditions and costs are given in Table 5.3.

Table 5.3 Conditions and cost of heating and cooling utilities.

Utility Name	Type	Temperature (° F)		Pressure (psia)	Cost
		in	**out**		
Low pressure steam	**Heating**	**330**	**329**	**70**	**$4.4 / MMBtu**
High pressure steam	**Heating**	**411**	**410**	**235**	**$6.0 / MMBtu**
Water	**Cooling**	**80**	**90**	**1.0**	**$0.04 / ton**
Brine	**Cooling**	**40**	**57**	**1.0**	**$0.08 / ton**
Propane	**Cooling**	**-40**	**-39**	**1.0**	**$ 20.0/MMBtu**
Ethylene	**Cooling**	**-150**	**-149**	**1.0**	**$30.0/ MMBtu**
Mixed refrigerant	**Cooling**	**-270**	**-269**	**1.0**	**$50.0/ MMBtu**
Electricity	**-**	**-**	**-**	**-**	**$0.045/kWh**
Gas	**-**	**-**	**-**	**-**	**$7.0/ MMBtu**

6. RESULTS AND DISCUSSIONS

This section provides the key results of the developed approach as applied to an LNG process.

6.1 Process Synthesis for Generation of Alternative Fractionation Configurations

First, a typical LNG flowsheet is synthesized and process alternatives are developed. The LNG flowsheet consists of mainly five units, the inlet receiving unit, acid treating unit, dehydration unit, fractionation unit, and liquefaction unit. At the inlet receiving unit, the separation between the gas and the condensate (heavy hydrocarbons) takes place. The slug catcher and the stabilizer are used mainly for that separation. For the gas treating unit the solvent absorption process is used to reduce the CO_2 and H_2S concentrations. The methyldiethanolamine (MDEA) is used as an aqueous solvent with 50 wt%. For the dehydration unit, the adsorption process is considered. Molecular sieve is used as adsorbent to remove water. For the fractionation unit, four distillation columns are used to recover various cuts of hydrocarbons. They are de-methanizer, de-ethanizer, de-proponizer, and du-butanizer columns. Finally, for the liquefaction unit the mixed refrigerant is used to cool the gas stream down to -256 ℉. Another task in this step is to develop some alternatives particularly in the fractionation unit for example the column sequences. Should the methane and ethane get separated first from the other hydrocarbons or just the methane then ethane and so on? Figure 6.1 shows the developed alternatives.

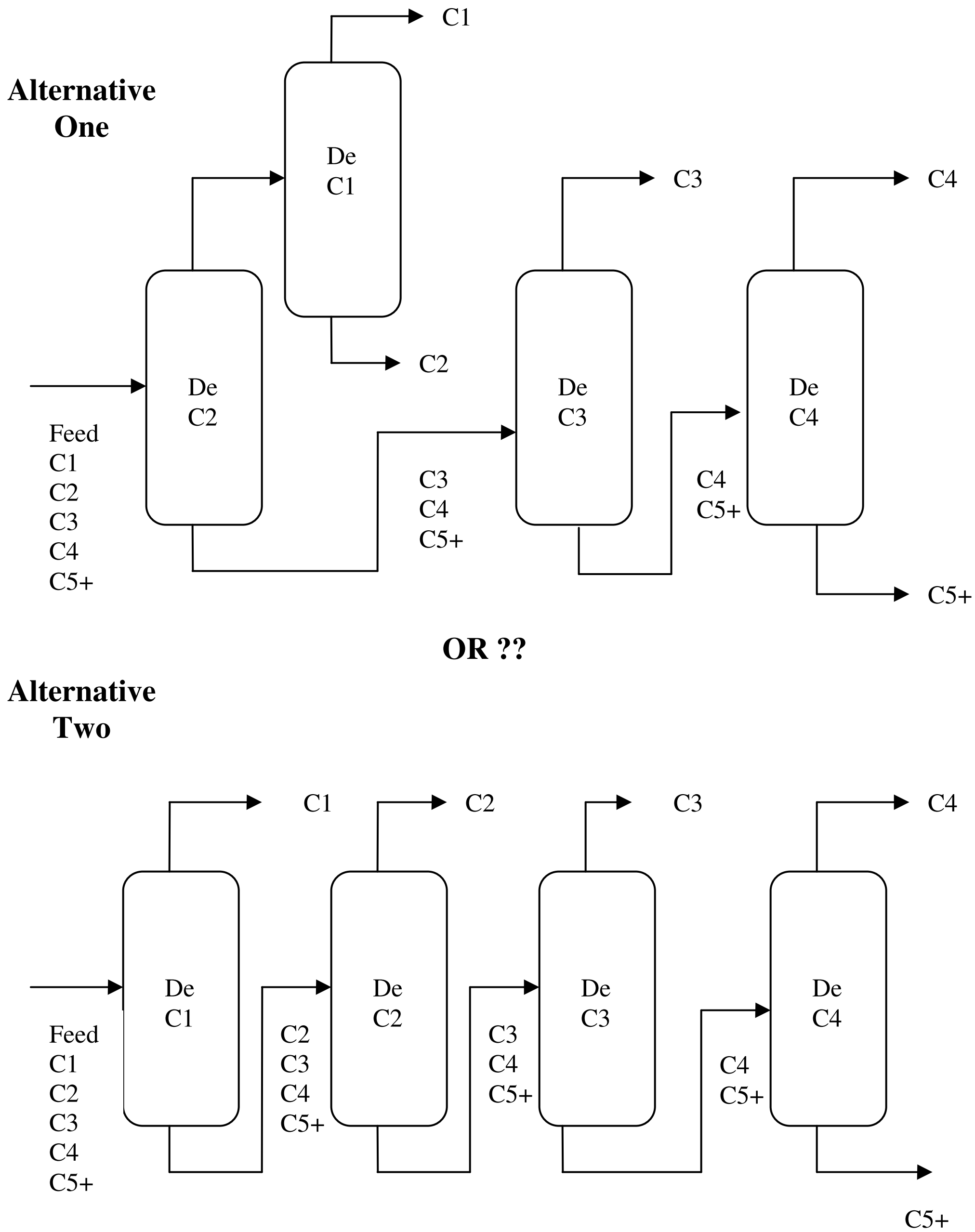

Figure 6.1 Two candidate alternatives for the fractionation train.

6.2 Simulation and Optimization Step Results

All the above units are simulated using ASPEN Plus. An appropriate thermodynamic property method (e.g., Peng Robinson) is selected as global flowsheet method. However, ELECNRTL is used in acid gas unit as block property method for better results and convergence. Since MDEA will be used as solvent in acid gas unit, KEMDEA (property package) is imported from ASPEN Plus data base. Now, the different process alternatives will be examined to select the superior configuration from energy usage, and columns sizes points of view. Table 6.1 and 6.2 show the results of alternative one and two respectively for feed precooling to -40 ℉ and fixed recovery of methane (99%) and ethane (75%).

Table 6.1 Alternative one results.

	DeC1	DeC2	DeC3	DeC4
Q_C, MMBtu /h	93	68	15	5
Overhead Temp, ℉	-191	-144	20	163
Q_R, MMBtu /h	25	31	15	5
Bottom temp, ℉	-29	54	181	232
No. of stages	12	20	25	25

Table 6.2 Alternative two results.

	DeC1	DeC2	DeC3	DeC4
Q_C, MMBtu /h	**111**	**71**	**7**	**3**
Overhead Temp, °F	**-164**	**-8**	**82**	**118**
Q_R, MMBtu /h	**33**	**72**	**7**	**3**
Bottom temp, °F	**14**	**115**	**196**	**216**
No. of stages	**20**	**30**	**25**	**25**

As can be seen from Table 6.2 the overhead temperature of DeC2 is -8 °F in alternative two where brine could be used to achieve this cooling. However, the overhead temperature of DeC2 is -144 °F in alternative one where a refrigerant should be used to achieve this cooling. Obviously, using the brine should lead to a significant saving in operating cost. Comparing other key factors for example, column sizes, condenser and boiler duties, alternative two is selected as a winner configuration. Figure 6.2 shows the process flowsheet for 1,500 MMSCFD. Based on the simulation results, the current heating utility is 2,441 MMBtu/h and the current cooling utility is 1,246 MMBtu/h. Figure 6.3 illustrates the process flowsheet with symbols and relevant data for the various heat exchangers.

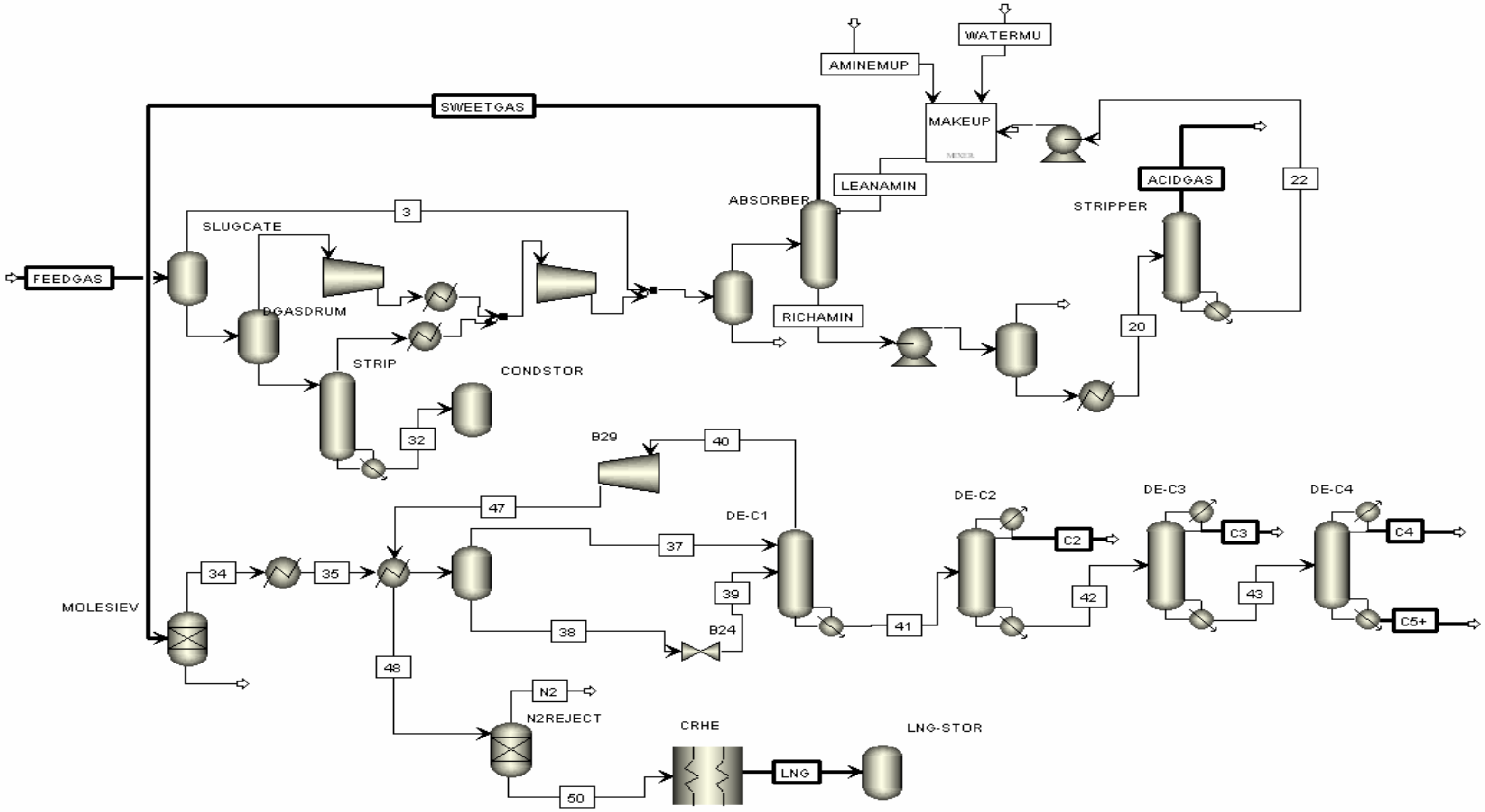

Figure 6.2 The flowsheet of the base-case LNG process.

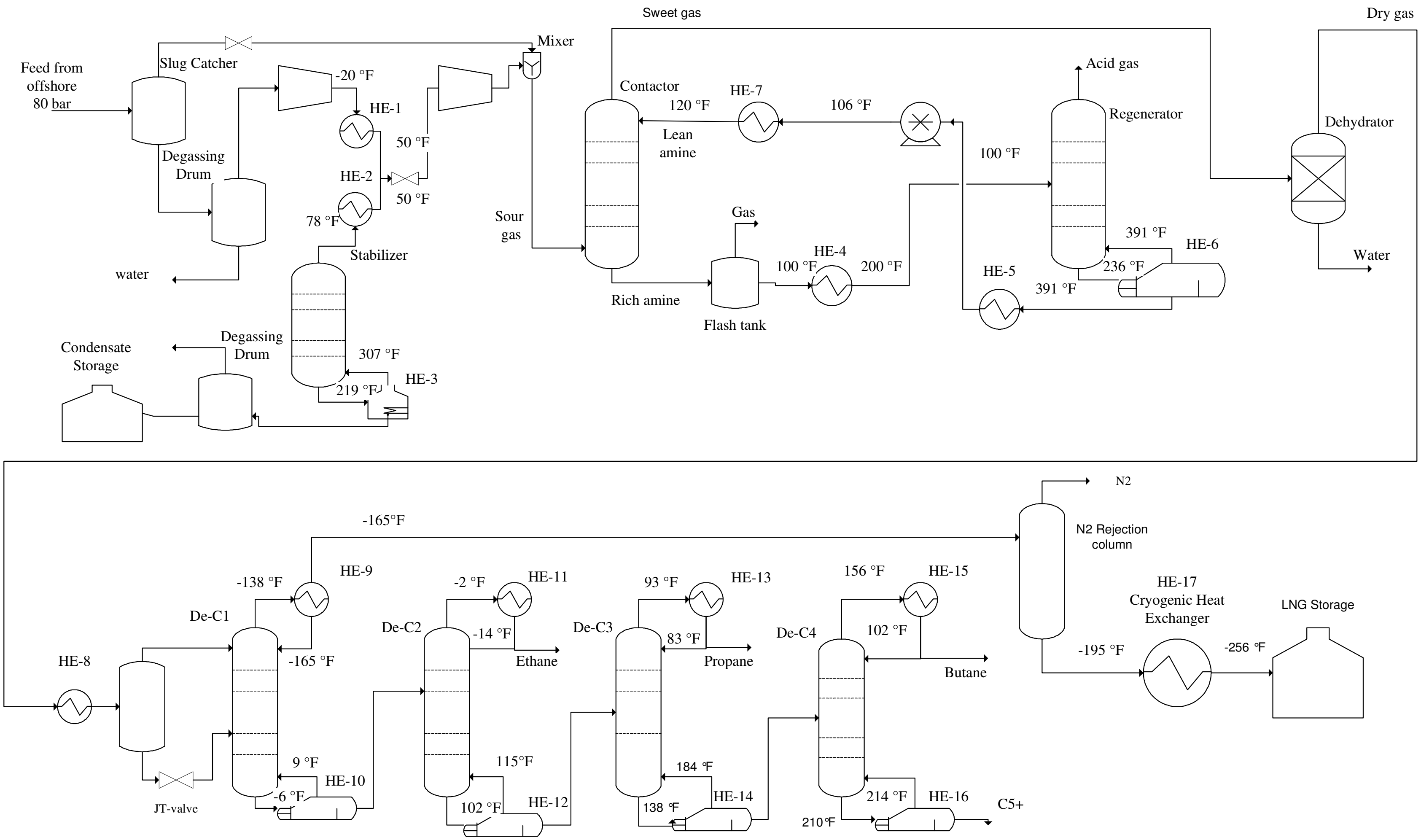

Figure 6.3 LNG process flowsheet with symbols of heat exchangers and temperatures.

6.3 Process Integration and Targeting Step Results

At this step, the process integration and targeting are applied to benchmark the performance of the process. Data extraction is carried out for the relevant temperatures and heat duties of the flowsheet described by Figure 6.3. Table 6.3 provides the supply and target temperatures of the process hot and cold streams as well as the utilities. The thermal pinch analysis is conducted to benchmark the potential of exchanging heat among the process hot and cold streams. The first step in algebraically developing the thermal pinch analysis is the construction of the temperature interval diagram as shown in Figure 6.4. A minimum approach temperature of 5 ℉ is used.

Next, the table of exchangeable heat loads (TEHL) for the process hot and cold streams is constructed. Having determined the individual loads of all process streams for all temperature intervals, one can calculate the collective loads (capacities) of the hot (cold) process streams.

Table 6.3 Extracted stream data for the LNG process from simulated flowsheet.

Description (exchanger or utility)	Stream Number	Flowrate*specific heat MMBtu/(h °F)	Supply Temp (°F)	Target Temp (°F)	Enthalpy change MMBtu/h
HE-5	**H1**	**1.15**	**391**	**100**	**-334.65**
HE-8	**H2**	**1.61**	**106**	**-35**	**-227.01**
HE-9	**H3**	**4.15**	**-138**	**-165**	**-112.05**
HE-11	**H4**	**6.10**	**-2**	**-14**	**-73.20**
HE-13	**H5**	**0.72**	**93**	**83**	**-7.20**
HE-15	**H6**	**0.06**	**156**	**102**	**-3.24**
HE-17	**H7**	**5.37**	**-165**	**-256**	**-488.67**
HP Steam	**H9(HU1)**	**?**	**411**	**410**	**?**
LP Steam	**H8(HU2)**	**?**	**330**	**329**	**?**
HE-3	**C1**	**0.56**	**219**	**307**	**49.28**
HE-4	**C2**	**4.08**	**100**	**200**	**408.00**
HE-6	**C3**	**12.08**	**237**	**391**	**1860.32**
HE-7	**C4**	**0.99**	**106**	**120**	**13.86**
HE-10	**C5**	**1.65**	**-6**	**9**	**24.75**
HE-12	**C6**	**5.77**	**102**	**115**	**75.01**
HE-14	**C7**	**0.15**	**138**	**184**	**6.90**
HE-16	**C8**	**0.75**	**210**	**214**	**3.00**
Cooling water	**C9(CU1)**	**?**	**78**	**88**	**?**
Brine water	**C10(CU2)**	**?**	**40**	**57**	**?**
Propane	**C11(CU3)**	**?**	**-40**	**-39**	**?**
Ethylene	**C12(CU4)**	**?**	**-150**	**-149**	**?**
Mixed refrigerant	**C13(CU5)**	**?**	**-270**	**-269**	**?**

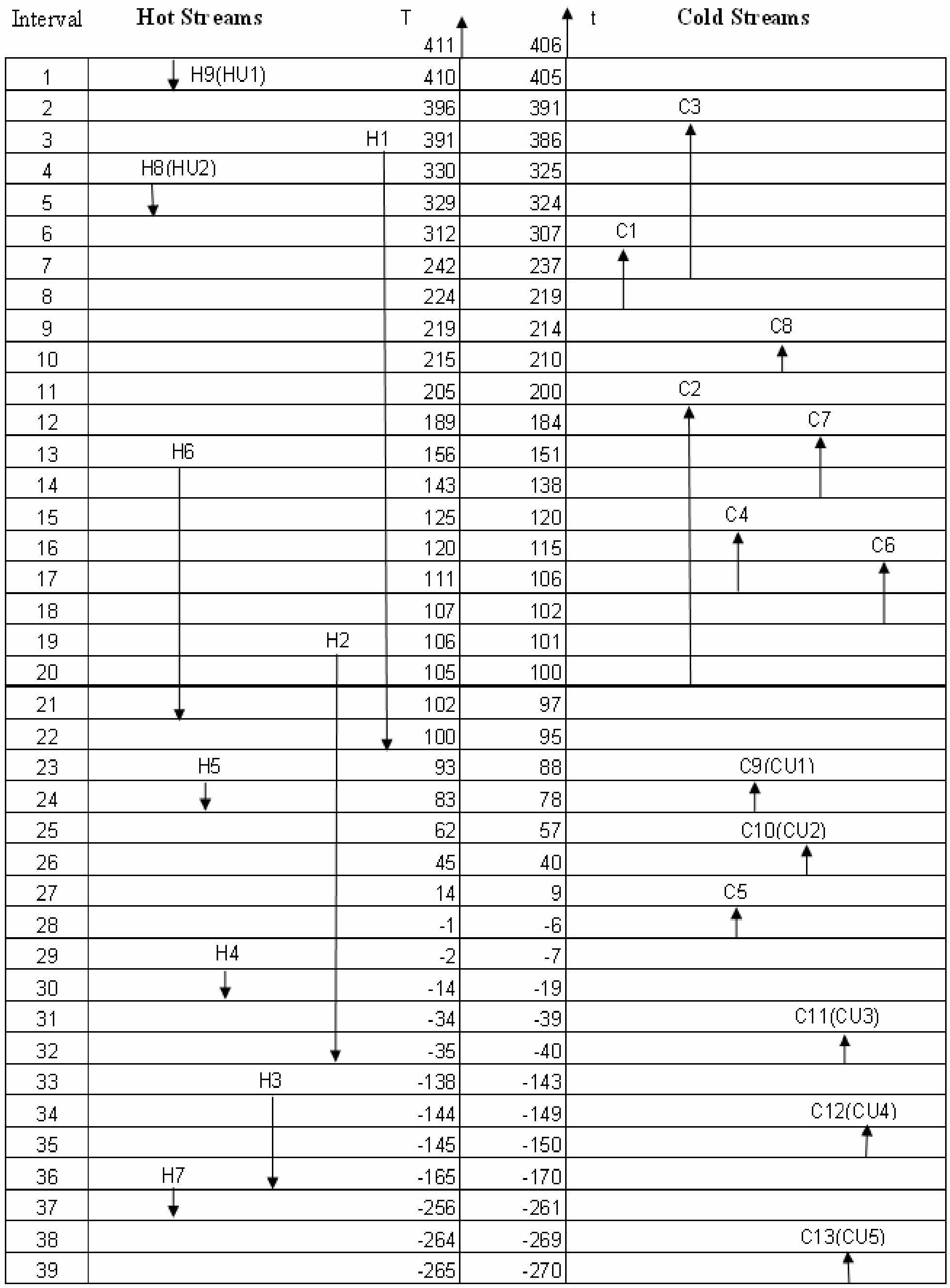

Figure 6.4 Temperature interval diagram for the LNG process.

Table 6.4 TEHL for process hot streams.

Interval	Load of H1 (MM Btu/h)	Load of H2 (MM Btu/h)	Load of H3 (MM Btu/h)	Load of H4 (MM Btu/h)	Load of H5 (MM Btu/h)	Load of H6 (MM Btu/h)	Load of H7 (MM Btu/h)	Total Load (MM Btu/h)
1	-	-	-	-	-	-	-	-
2	-	-	-	-	-	-	-	-
3	-	-	-	-	-	-	-	-
4	70.15	-	-	-	-	-	-	70.15
5	1.15	-	-	-	-	-	-	1.15
6	19.55	-	-	-	-	-	-	19.55
7	80.50	-	-	-	-	-	-	80.50
8	20.70	-	-	-	-	-	-	20.70
9	5.75	-	-	-	-	-	-	5.75
10	4.60	-	-	-	-	-	-	4.60
11	11.50	-	-	-	-	-	-	11.50
12	18.40	-	-	-	-	-	-	18.40
13	37.95	-	-	-	-	-	-	37.95
14	14.95	-	-	-	-	0.78	-	15.73
15	20.70	-	-	-	-	1.08	-	21.78
16	5.75	-	-	-	-	0.30	-	6.05
17	10.35	-	-	-	-	0.54	-	10.89
18	4.60	-	-	-	-	0.24	-	4.84
19	1.15	-	-	-	-	0.06	-	1.21
20	1.15	1.61	-	-	-	0.06	-	2.82
21	3.45	4.83	-	-	-	0.18	-	8.46
22	2.30	3.22	-	-	-	-	-	5.52
23	-	11.27	-	-	-	-	-	11.27
24	-	16.10	-	-	7.20	-	-	23.3
25	-	33.81	-	-	-	-	-	33.81
26	-	27.37	-	-	-	-	-	27.37
27	-	49.91	-	-	-	-	-	49.91
28	-	24.15	-	-	-	-	-	24.15
29	-	1.61	-	-	-	-	-	1.61
30	-	19.32	-	73.20	-	-	-	92.52
31	-	32.20	-	-	-	-	-	32.20
32	-	1.61	-	-	-	-	-	1.61
33	-	-	-	-	-	-	-	-
34	-	-	24.90	-	-	-	-	24.90
35	-	-	4.15	-	-	-	-	4.15
36	-	-	83.00	-	-	-	-	83.00
37	-	-	-	-	-	-	488.67	488.67
38	-	-	-	-	-	-	-	-
39	-	-	-	-	-	-	-	-

Table 6.5 TEHL for process cold stream.

Interval	Load of C1 (MM Btu/h)	Load of C2 (MM Btu/h)	Load of C3 (MM Btu/h)	Load of C4 (MM Btu/h)	Load of C5 (MM Btu/h)	Load of C6 (MM Btu/h)	Load of C7 (MM Btu/h)	Load of C8 (MM Btu/h)	Total Load (MM Btu/h)
1	-	-	-	-	-	-	-	-	-
2	-	-	-	-	-	-	-	-	-
3	-	-	60.40	-	-	-	-	-	60.40
4	-	-	736.88	-	-	-	-	-	736.88
5	-	-	12.08	-	-	-	-	-	12.08
6	-	-	205.36	-	-	-	-	-	205.36
7	39.20	-	845.60	-	-	-	-	-	884.80
8	10.08	-	-	-	-	-	-	-	10.08
9	-	-	-	-	-	-	-	-	-
10	-	-	-	-	-	-	-	3.00	3.00
11	-	-	-	-	-	-	-	-	-
12	-	65.28	-	-	-	-	-	-	65.28
13	-	134.64	-	-	-	-	4.95	-	139.59
14	-	53.04	-	-	-	-	1.95	-	54.99
15	-	73.44	-	-	-	-	-	-	73.44
16	-	20.40	-	4.95	-	-	-	-	25.35
17	-	36.72	-	8.91	-	51.93	-	-	97.56
18	-	16.32	-	-	-	23.08	-	-	39.40
19	-	4.08	-	-	-	-	-	-	4.08
20	-	4.08	-	-	-	-	-	-	4.08
21	-	-	-	-	-	-	-	-	-
22	-	-	-	-	-	-	-	-	-
23	-	-	-	-	-	-	-	-	-
24	-	-	-	-	-	-	-	-	-
25	-	-	-	-	-	-	-	-	-
26	-	-	-	-	-	-	-	-	-
27	-	-	-	-	-	-	-	-	-
28	-	-	-	-	24.75	-	-	-	24.75
29	-	-	-	-	-	-	-	-	-
30	-	-	-	-	-	-	-	-	-
31	-	-	-	-	-	-	-	-	-
32	-	-	-	-	-	-	-	-	-
33	-	-	-	-	-	-	-	-	-
34	-	-	-	-	-	-	-	-	-
35	-	-	-	-	-	-	-	-	-
36	-	-	-	-	-	-	-	-	-
37	-	-	-	-	-	-	-	-	-
38	-	-	-	-	-	-	-	-	-
39	-	-	-	-	-	-	-	-	-

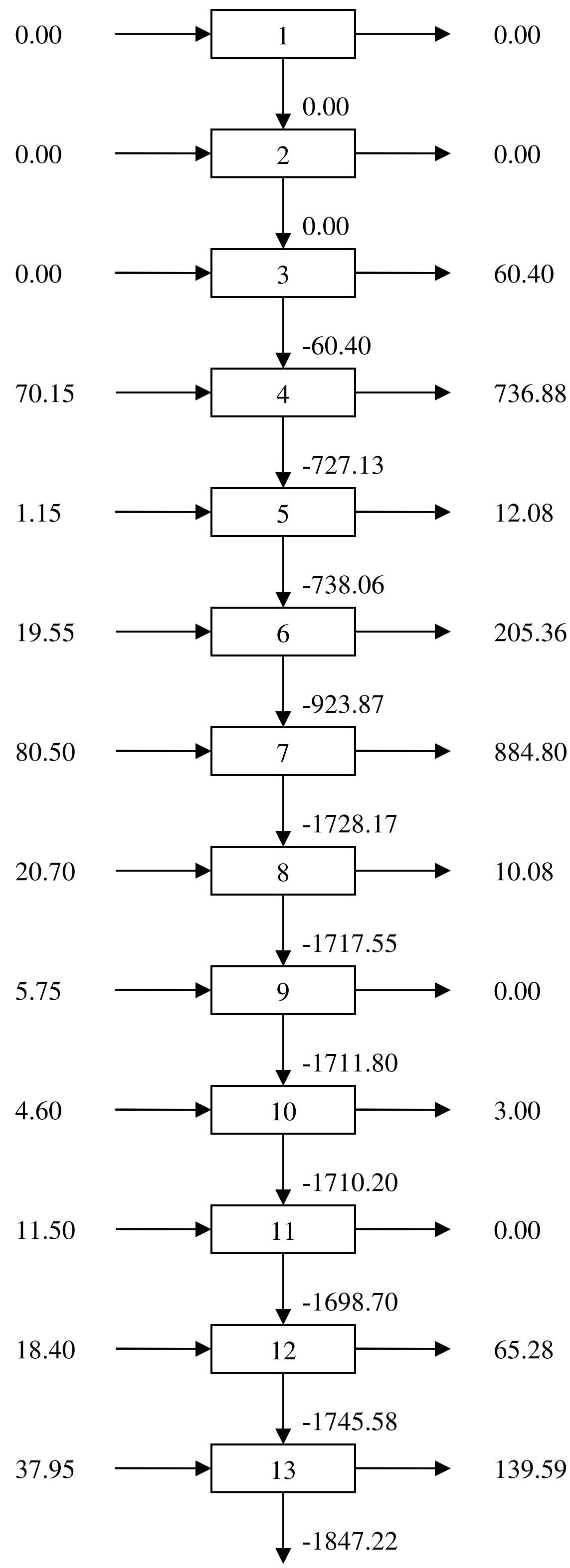

Figure 6.5 Cascade diagram for the LNG process.

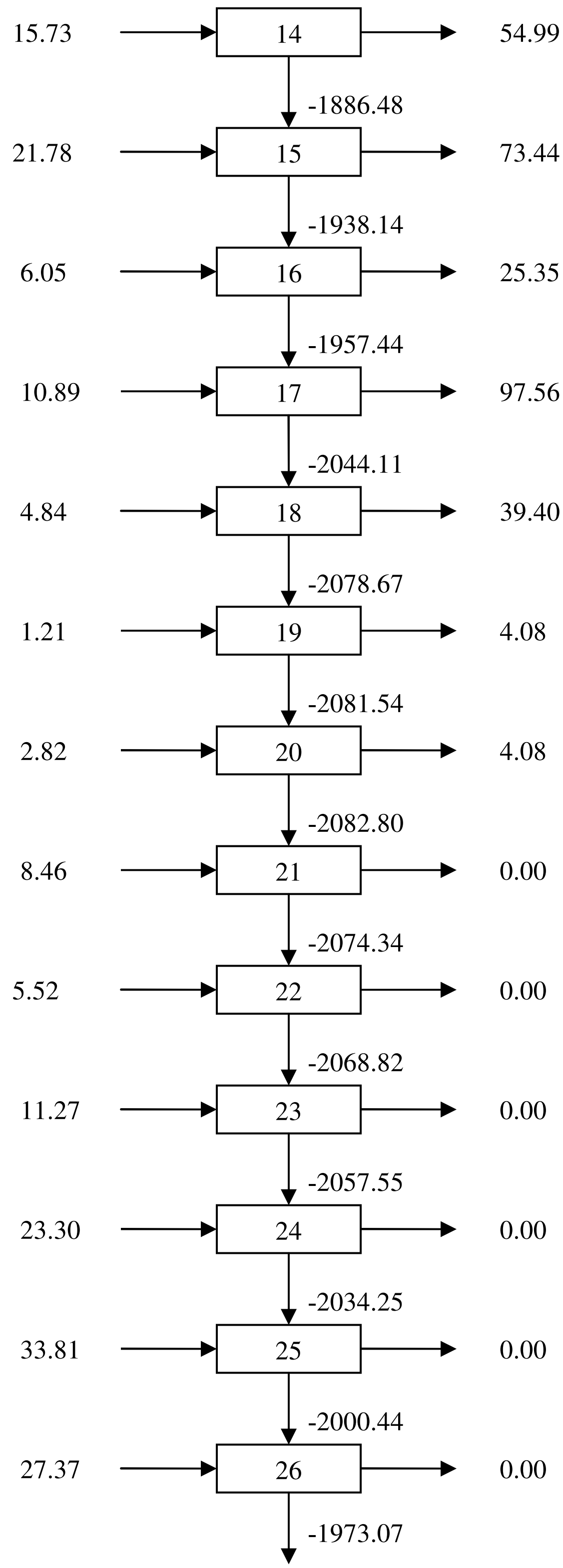

Figure 6.5 Continued

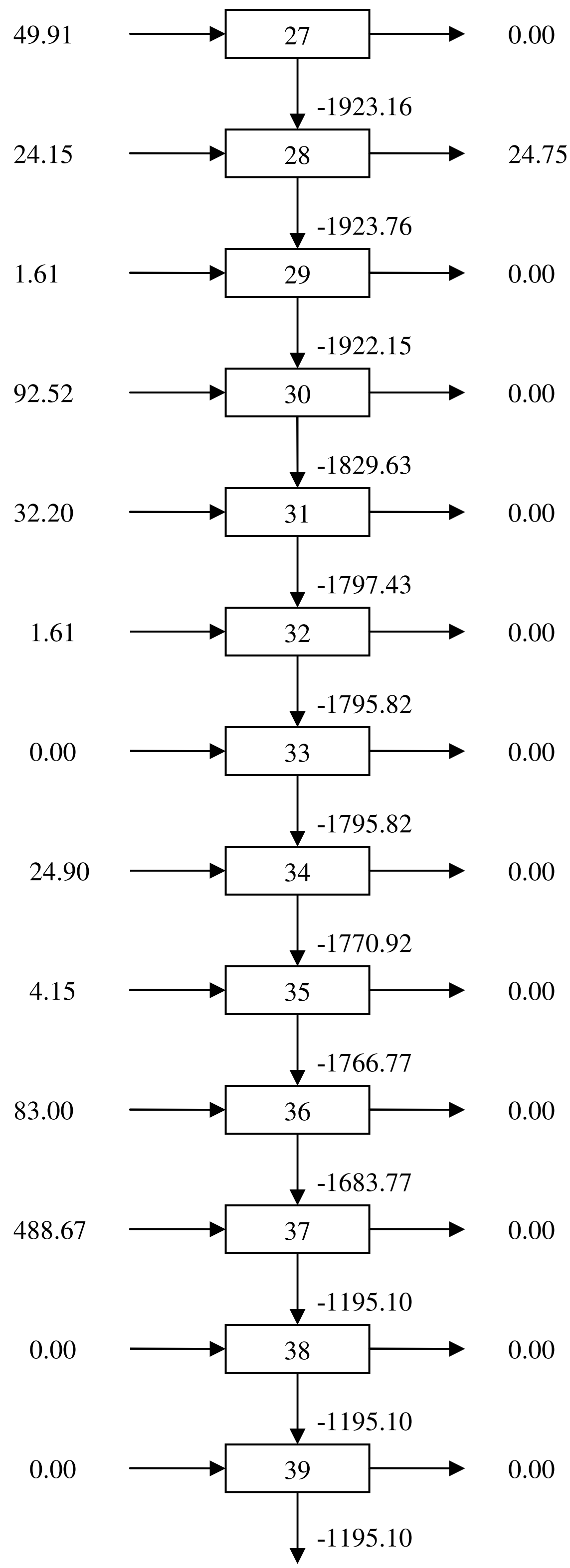

Figure 6.5 Continued

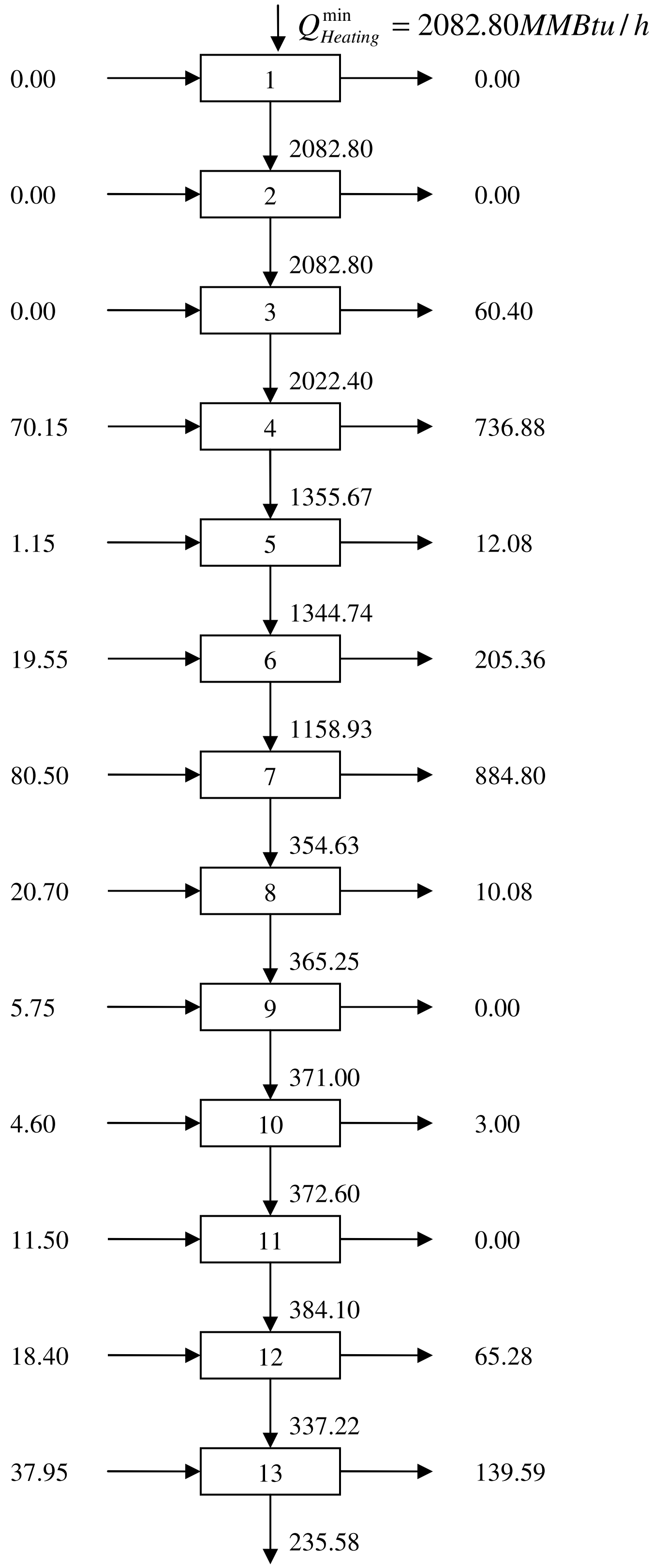

Figure 6.6 Revised cascade diagram for the LNG process.

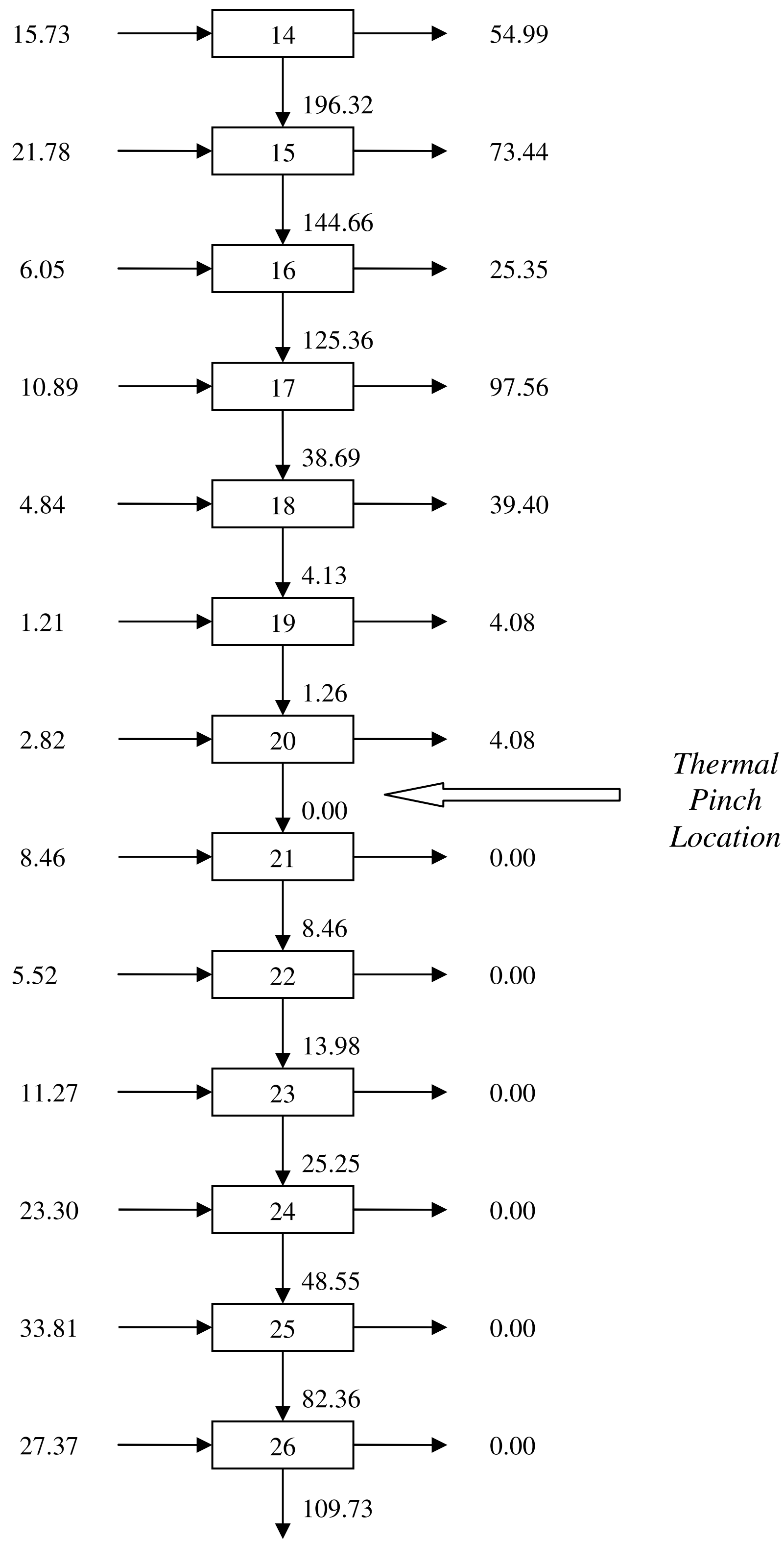

Figure 6.6 Continued

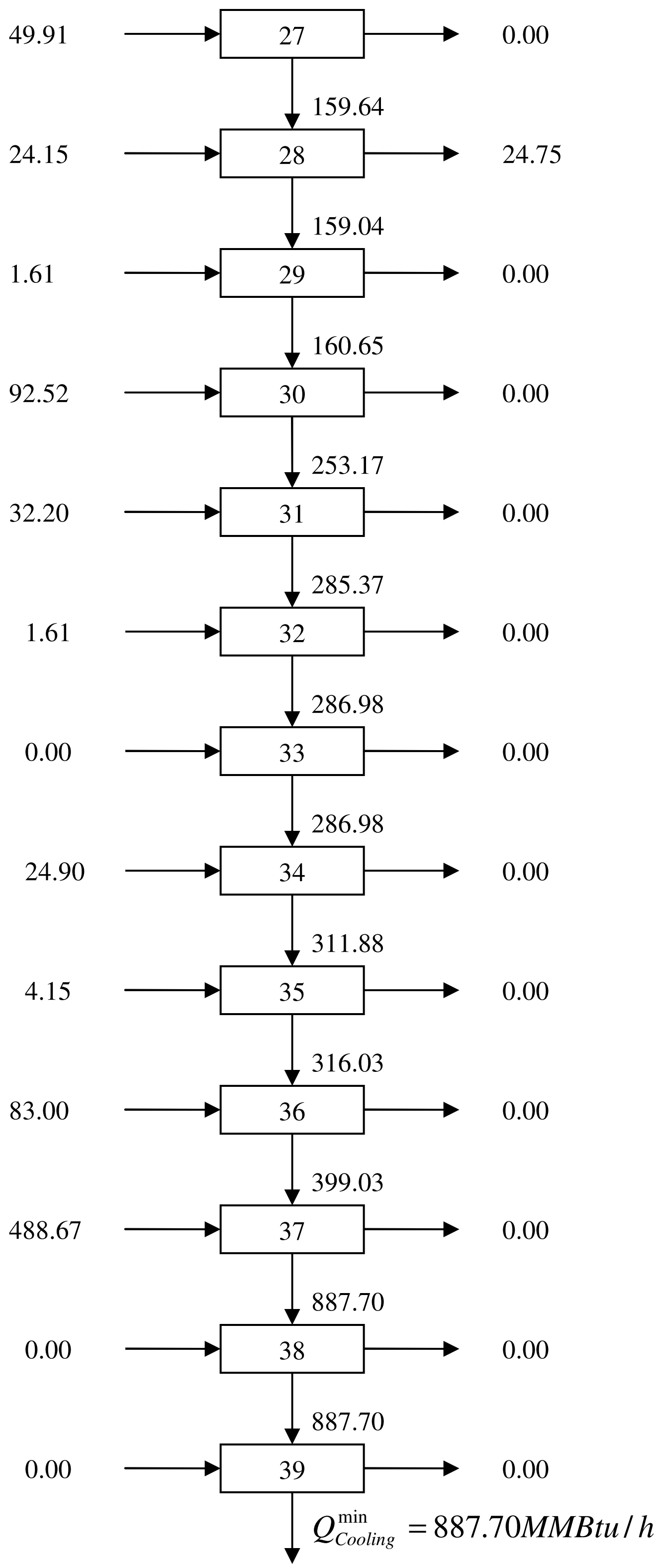

Figure 6.6 Continued

Table 6.4 and 6.5 show TEHL for process hot and cold streams, respectively. The cascade diagram calculations are shown in Figure 6.5. By adding the most negative value from cascade diagram to the top interval in the revised cascade diagram as a positive value, the minimum utility requirements are obtained as shown in Figure 6.6. The minimum heating utility ($Q_{Heating}^{\min}$) is 2082.80 MMBtu/h and the minimum cooling utility ($Q_{Cooling}^{\min}$) is 887.70 MMBtu/h. Also, the thermal pinch point is located between intervals 20 and 21 which represents 105 °F on the hot scale (100 °F on the cold stream). Therefore, the reduction in utilities can be calculated as follows:

Target for percentage savings in heating utility= $\dfrac{2441.12 - 2082.80}{2441.12} \times 100\% = 15\%$

Target for percentage savings in cooling utility= $\dfrac{1246.02 - 887.70}{1246.02} \times 100\% = 29\%$

Now, many utilities are available for service. These utilities can be screened to minimize the operating cost. A convenient way of screening multiple utilities is by constructing the grand composite curve (GCC). Figure 6.7 shows GCC for the LNG process.

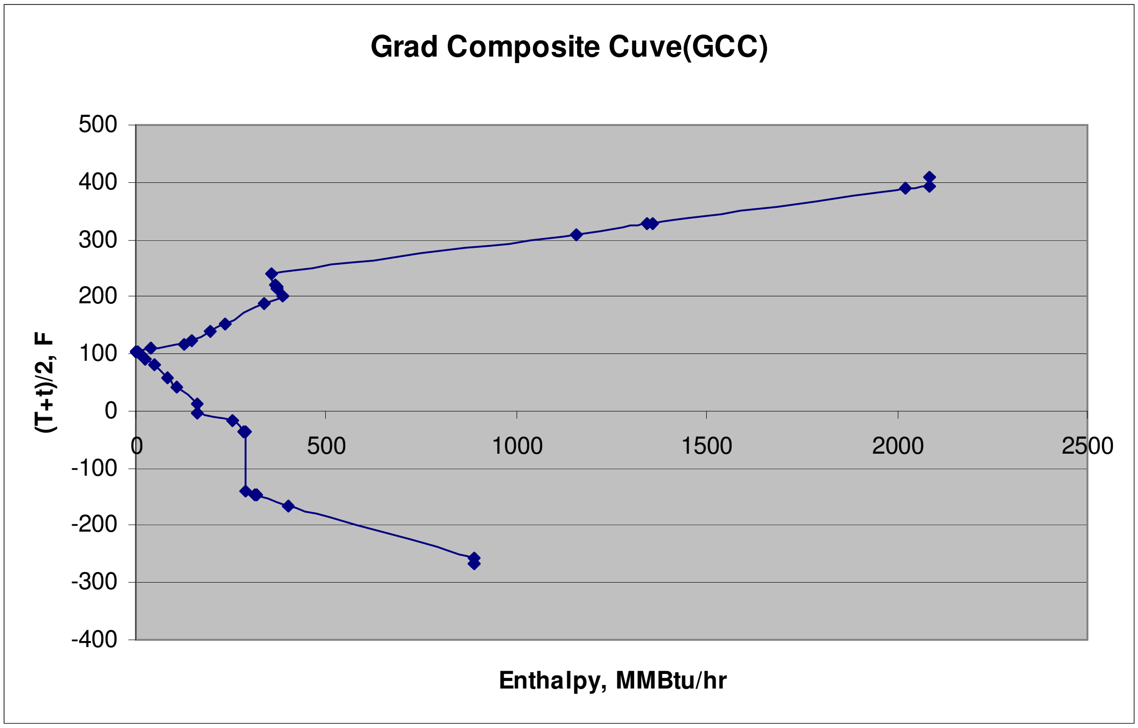

Figure 6.7 Grand composite curve for the LNG process.

The actual heat transfer areas of the heat exchangers are obtained from simulation using ASPEN Plus. Figure 6.8 shows the title of each heat exchanger and its actual heat transfer area. For practicality reasons, the retrofitting analysis will consider heat exchangers with heat transfer areas exceeding 500 ft^2. Therefore, only the following heat exchanger will be included in the retrofitting analysis: HE-3, HE-4, HE-5, HE-6, HE-8, HE-9, HE-13, and HE-17.

The pinch point divides the problem into two parts; above and below the pinch. A mixed integer linear program (MILP) is written and solved in LINGO. It aims to find the optimal matching between the streams while attaining the minimum utility requirements. Appendix A1 gives the formulation of the problem by using the generic formulation code of (Papoulias and Grossman 1979) that was described in section 2. The global optimal solution was obtained and appendix A2 gives all the results obtained from the optimization software LINGO. The resulted heat exchangers are: E9_3_1, E8_3_1, E8_1_1, E8_8_1, E8_2_1, E8_4_1, E8_6_1, E1_3_1, E1_2_1, E1_7_1, E2_10_2, E2_5_2, E2_11_2, E5_9_2, E3_13_2, and E7_13_2. By knowing the exchangeable loads within each heat exchanger from LINGO output, the inlet or outlet temperature of each stream are calculated and the network of heat exchangers is configured. Appendix B gives the relevant calculations for the network synthesis.

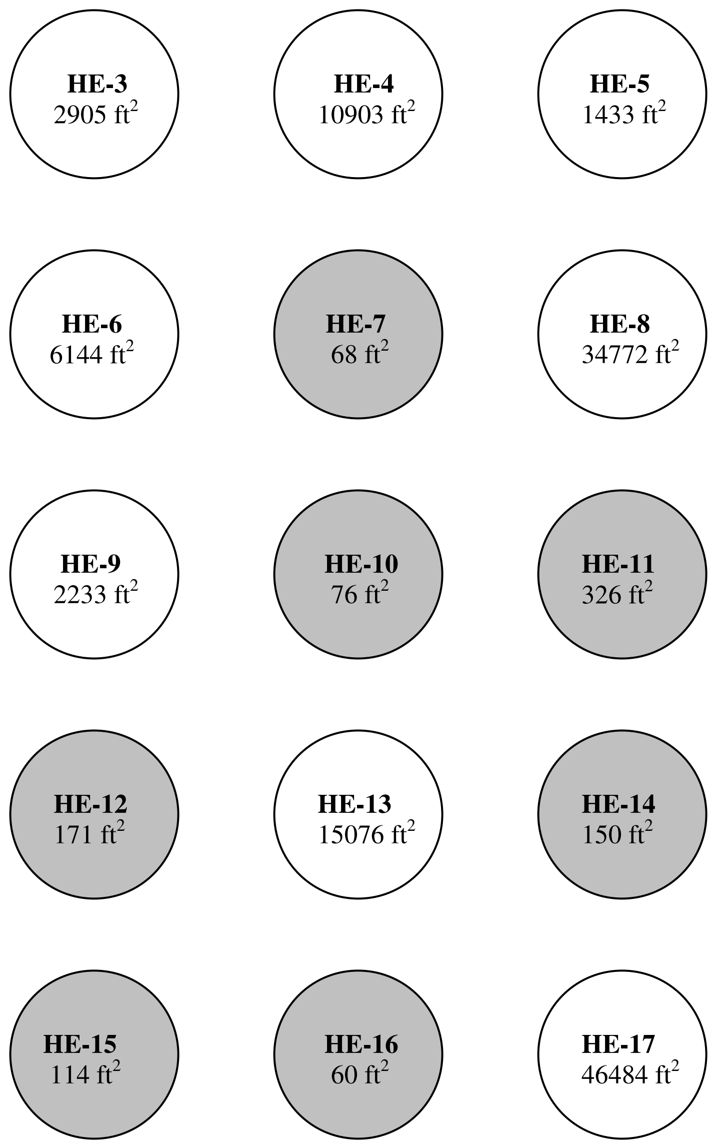

Figure 6.8 Actual heat transfer area of the simulated heat exchangers. (Shaded circles designate heat exchangers too small to be practically retrofitted).

6.4 Sizing and Design Step Results

A rigorous steady state simulation was performed for the integrated heat exchangers to evaluate their sizes and designs. Figure 6.9 shows the actual heat transfer area for the integrated heat exchangers. As mentioned earlier, for practicality reasons, any heat exchanger having heat transfer area less than 500 ft^2 will not be included in the retrofitting analysis. After conducting the rigorous steady state ASPEN simulation, the economic analysis step comes to evaluate the fixed and operating costs for every simulated exchanger.

6.5 Retrofitting Step Results

The retrofitting step is performed with three scenarios (cases) of objective functions. The first case takes the fixed charges of the new exchanger into account. Next, the fixed charges are not considered in the second case. Finally, the third case is the simplest case in which the objective function is aimed to minimize the heat transfer area of the new exchangers.

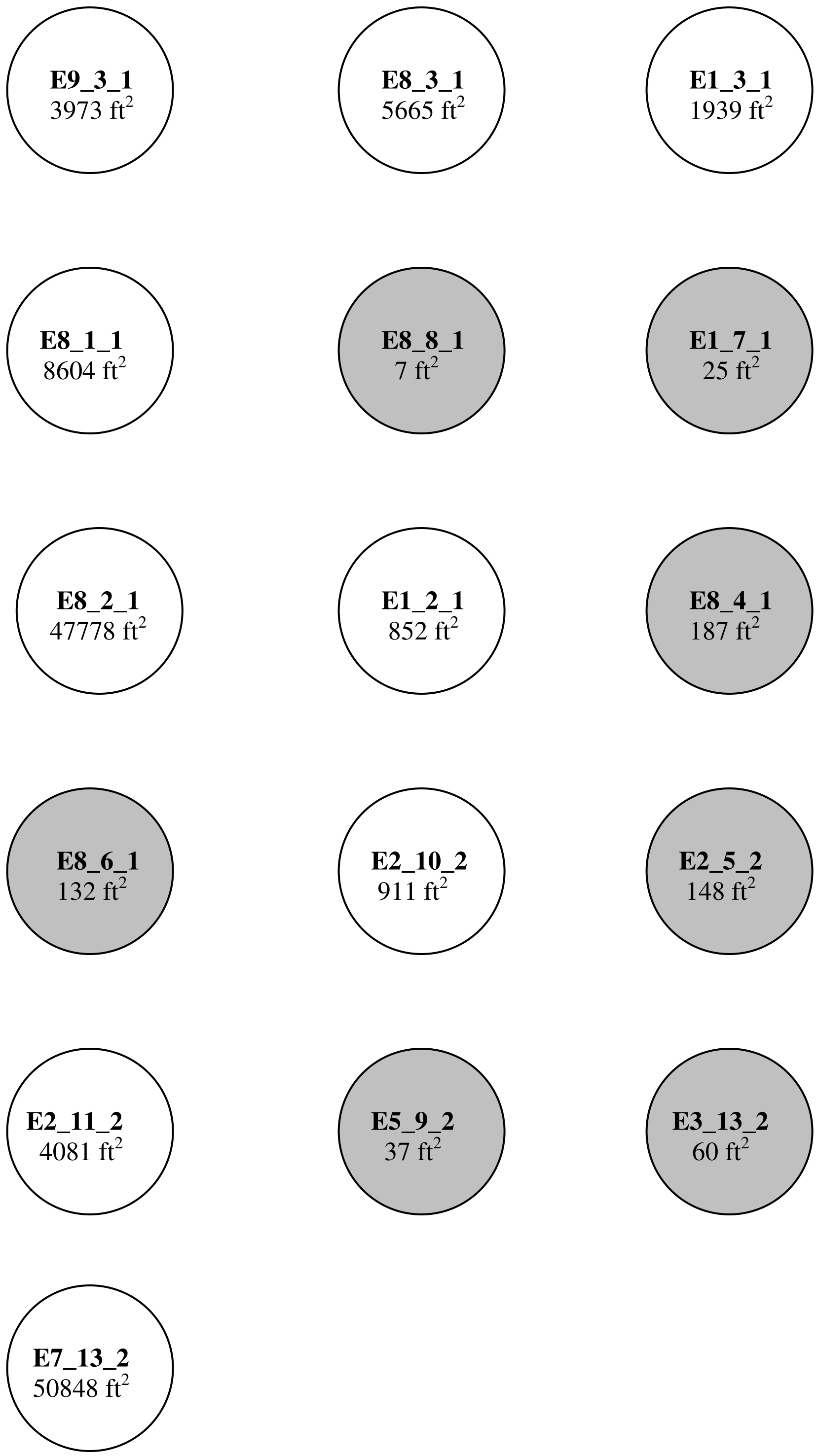

Figure 6.9 Actual heat transfer area of the integrated heat exchangers. (Shaded circles designate heat exchangers too small to be practically retrofitted).

The mathematical expressions for the three objective functions are given by:

Minimize $\sum_{i} (a_i + b_i * A_i^{new^x}) * E_i$

where a_i is a constant reflecting the fixed charges associated with the site-preparation work involved in the installation of the exchanger and the constants b_i and x (usually x ~0.6) are used to evaluate the fixed cost of the exchanger. E_i is a binary integer variable designating the presence or absence of the heat exchanger.

Therefore, the following new formulation is proposed:

If there are no fixed charges for each new exchanger, then the objective function is simplified to:

Minimize $\sum_{i} b_i * A_i^{new^x}$

Finally, if the objective is to minimize the heat transfer area of the new exchangers, then the objective function becomes the following linear expression:

Minimize $\sum_{i} A_i^{new}$

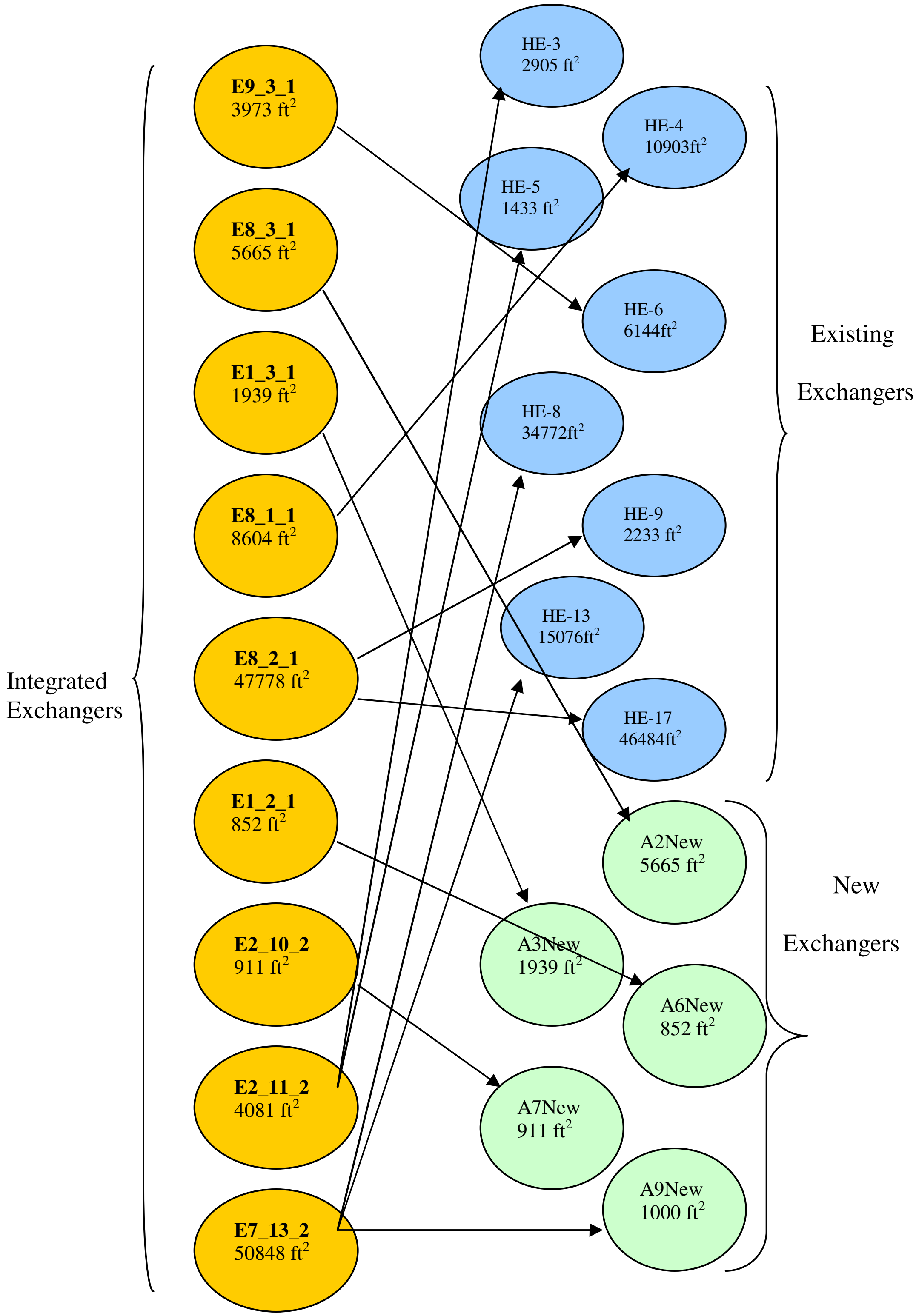

Figure 6.10 Retrofitting task with fixed charges consideration (1st case).

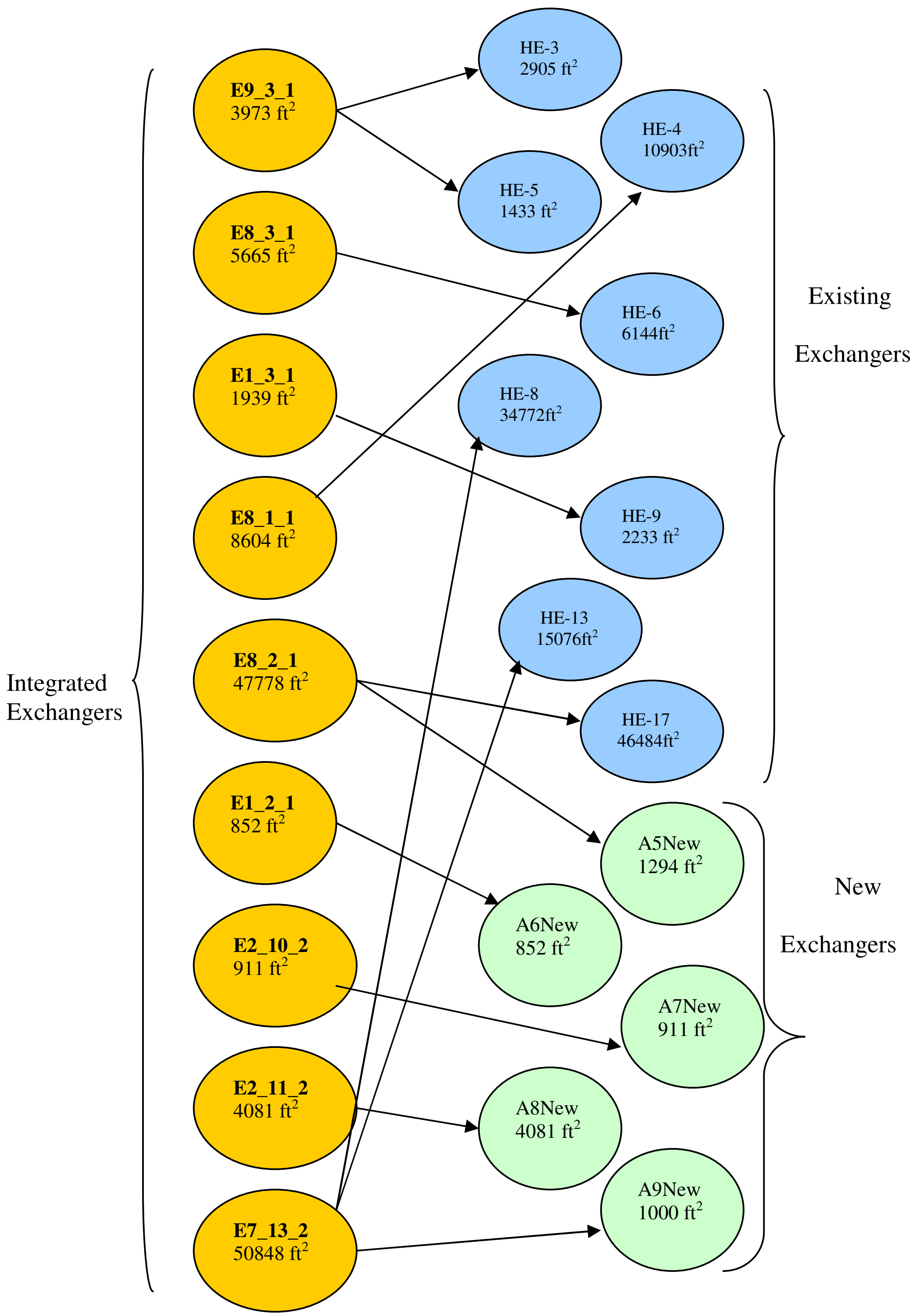

Figure 6.11 Retrofitting task without fixed charges consideration (2nd case).

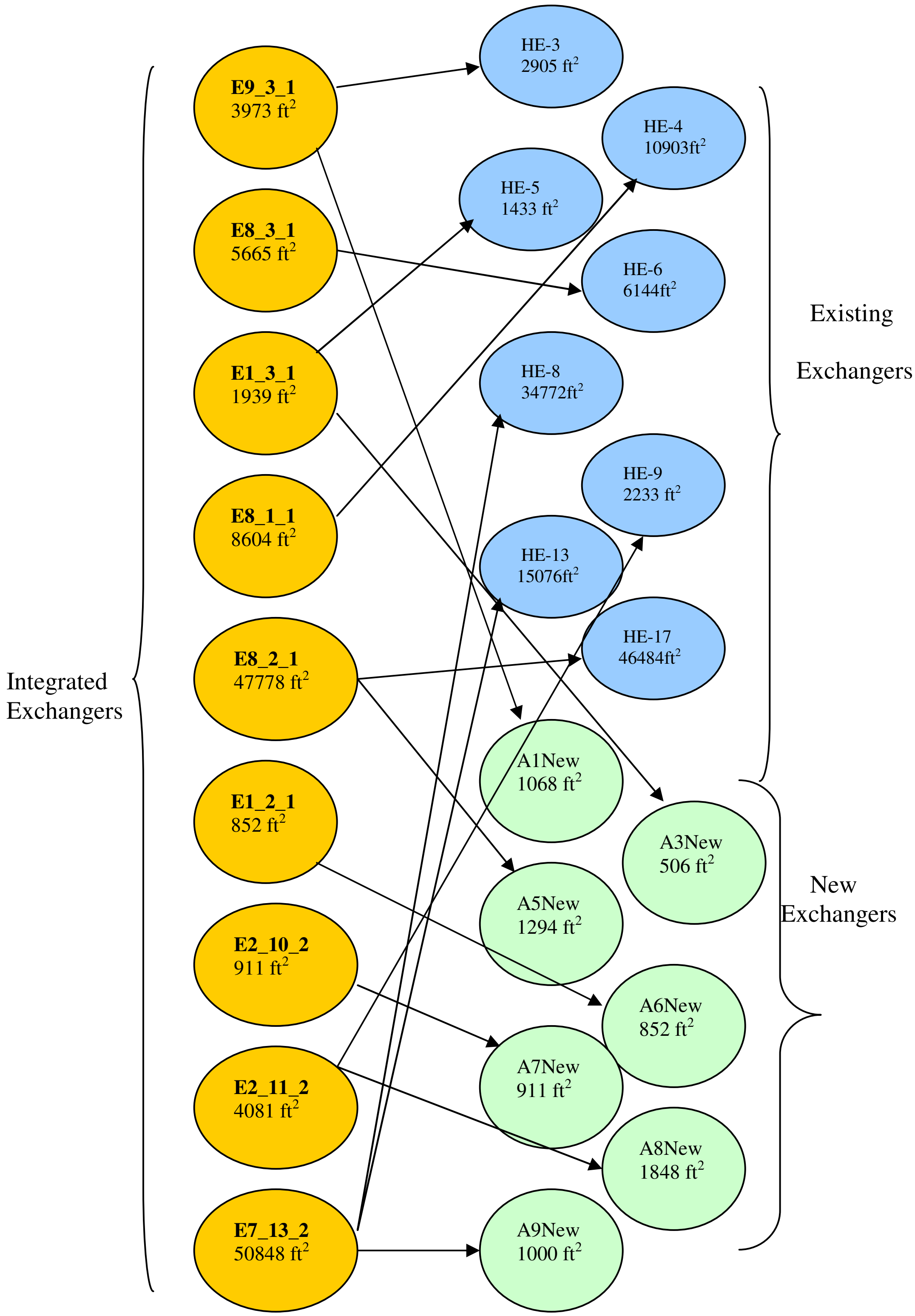

Figure 6.12 Retrofitting task for minimum heat transfer area (3rd case).

By formulating the retrofitting MINLPs described in section four and solving them using LINGO, the obtained new heat transfer areas are 10,367 ft^2, 8,138 ft^2, and 7,912 ft^2 for the three cases as shown in Figure 6.10, 6.11, and 6.12, respectively. The results from case one will be considered since its objective function consider the fixed charges which reflect the more realistic situation.

Considering the retrofitting of the first case (objective function is heat-exchanger fixed cost with fixed charges), five new heat exchangers must be purchased with a total area of 10,367 ft^2. While integrated heat exchange saves heating and cooling utilities, such savings must justify the cost of installing new heat exchangers. The payback period is used as an economic indicator to aid the cost/benefit analysis for installing a new heat exchanger.

The overall payback period of the five exchangers is calculated to know if the whole retrofitting project is worth it or not. The ICARUS fixed cost estimation for all new heat exchangers are given in the following tables.

For Exchanger E1_2_1:

The fixed cost estimation is done by using ICARUS and it is $ 145,610. The detailed fixed cost estimation is given in Table 6.6.

Table 6.6 ICARUS fixed cost estimation for exchanger E1_2_1.

Item	**Material(USD)**	**Manpower(USD)**	**Manhours**
Equipment&Setting	17400.	885.	40
Piping	18106.	10544.	457
Civil	72000.	1235.	71
Structural Steel	0.	0.	0
Instrumentation	7660.	2844.	125
Electrical	0.	0.	0
Insulation	9145.	4434.	224
Paint	398.	959.	59
Subtotal	**124,709**	**20,901**	**976**

For Exchanger E1_3_1:

The fixed cost estimation is done by using ICARUS and it is $ 128,786. The detailed estimation is given in Table 6.7.

Table 6.7 ICARUS fixed cost estimation for exchanger E1_3_1.

Item	**Material(USD)**	**Manpower(USD)**	**Manhours**
Equipment&Setting	14100.	885.	40
Piping	13629.	9308.	404
Civil	68700.	1192.	69
Structural Steel	0.	0.	0
Instrumentation	7660.	2844.	125
Electrical	0.	0.	0
Insulation	5798.	3646.	185
Paint	299.	725.	45
Subtotal	**110,186**	**18,600**	**868**

For Exchanger E7_13_2:

The fixed cost estimation is done by using ICARUS and it is $ 151,873. The detailed estimation is given in Table 6.8.

Table 6.8 ICARUS fixed cost estimation for exchanger E7_13_2.

Item	**Material(USD)**	**Manpower(USD)**	**Manhours**
Equipment&Setting	20800.	885.	40
Piping	9745.	17956.	778
Civil	60079.	1161.	67
Structural Steel	0.	0.	0
Instrumentation	13132.	2844.	125
Electrical	0.	0.	0
Insulation	11718.	13553.	697
Paint	0.	0.	0
Subtotal	**115,474**	**36,399**	**1707**

For Exchanger E2_10_2:

The fixed cost estimation is done by using ICARUS and it is $ 147,161. The detailed estimation is given in Table 6.9.

Table 6.9 ICARUS fixed cost estimation for exchanger E2_10_2.

Item	**Material(USD)**	**Manpower(USD)**	**Manhours**
Equipment&Setting	26400.	885.	40
Piping	22761.	10840.	469
Civil	61079.	1161.	67
Structural Steel	0.	0.	0
Instrumentation	7705.	2844.	125
Electrical	0.	0.	0
Insulation	7895.	4234.	214
Paint	398.	959.	59
Subtotal	**123,238**	**20,923**	**974**

Table 6.10 ICARUS fixed cost estimation for exchanger E8_3_1.

Item	**Material(USD)**	**Manpower(USD)**	**Manhours**
Equipment&Setting	59100.	902.	41
Piping	31505.	14298.	620
Civil	67086.	1710.	98
Structural Steel	0.	0.	0
Instrumentation	8730.	2853.	125
Electrical	0.	0.	0
Insulation	16290.	8655.	439
Paint	601.	1441.	89
Subtotal	**213,171**	**29,859**	**1412**

For Exchanger E8-3-1:

The fixed cost estimation is done by using ICARUS and it is \$ 213,171. The detailed estimation is given in Table 6.10.

Therefore, the total investment of all newly installed heat exchangers is \$ 786,601.

- Savings in heating utility:

$$\frac{358.32MMBtu}{h} \times \frac{\$6}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$18,833,299}{yr}$$

- Savings in cooling utility:

$$\frac{358.32MMBtu}{h} \times \frac{\$7}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$21,972,182}{yr}$$

- Total annual savings = \$ 18,833,299 + \$ 21,972,182 = \$ 40,805,481

Payback period = $\dfrac{\$786,601}{\dfrac{\$40,805,481}{yr}} = 0.02\,yr$

This is clearly a very attractive payback and, therefore, justifies the overall retrofitting activities.

The pay back period for two out of five exchangers that utilized by process streams namely E1_2_1 and E1_3_1 is particularly calculated to validate the retrofitting activity. They have heat transfer area of 1,939 ft^2 and 852 ft^2, respectively.

For Exchanger E1_2_1:

- Savings in heating utility:

$$\frac{80MMBtu}{h} \times \frac{\$6}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$4,204,800}{yr}$$

- Savings in cooling utility:

$$\frac{80MMBtu}{h} \times \frac{\$7}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$4,905,600}{yr}$$

- Total annual savings = \$ 4,204,800+ \$ 4,905,600= \$ 9,110,400

Payback period = $\dfrac{\$145,610}{\dfrac{\$9,110,400}{yr}} = 0.02\,yr$

This is clearly a very attractive payback and, therefore, justifies the retrofitting activities.

For Exchanger E1_3_1:

- Savings in heating utility:

$$\frac{133MMBtu}{h} \times \frac{\$6}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$6{,}990{,}480}{yr}$$

- Savings in cooling utility:

$$\frac{133MMBtu}{h} \times \frac{\$7}{MMBtu} \times \frac{24h}{day} \times \frac{365day}{yr} = \frac{\$8{,}155{,}560}{yr}$$

- Total annual savings = \$ 6,990,480+\$ 8,155,560= \$ 15,146,000

Payback period = $\dfrac{\$128{,}786}{\dfrac{\$15{,}146{,}000}{yr}} = 0.01yr$

Again, this is a very attractive payback period indicating the worth of the retrofitting activities.

6.6 Turbo-expansion Activity Results

Here, the turbo-expansion activity is addressed in the fractionation unit. Currently, the feed gas stream is cooled and its pressure is reduced by a Joule-Thompson (JT) valve as shown in Figure 6.13. The pressure is reduced from 870 psia to 250 psia at which the de-methanizer column gives the desired recovery (e.g., 99% of methane in the top product and 75% of ethane in the bottom product). The condenser duty is 113 MMBtu/h and the reboiler duty is 25 MMBtu/h. The temperatures of the top and the bottom products are -165 °F and 9 °F, respectively

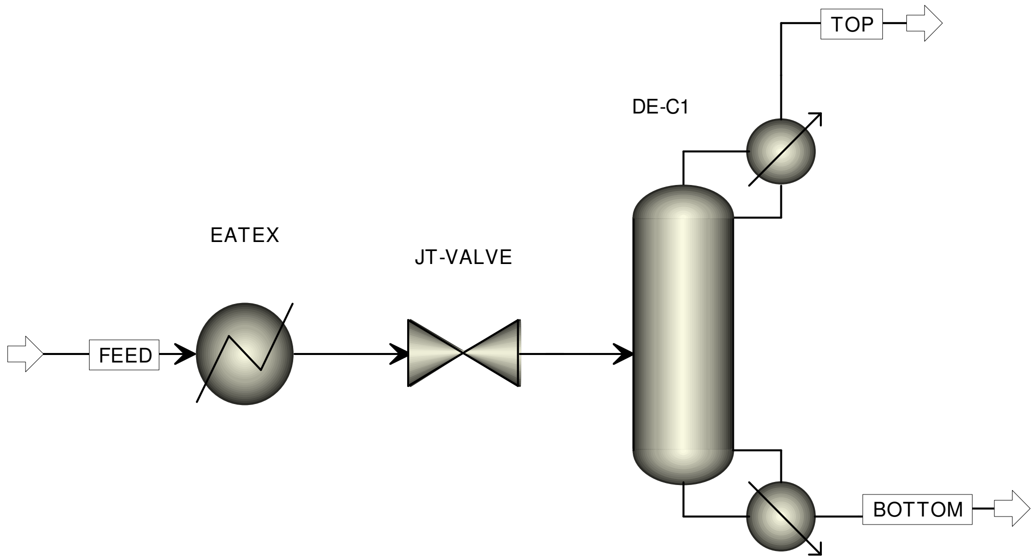

Figure 6.13 Current configuration of the de-methanizer column.

As can be seen from Figure 6.14, the expander is used instead of JT valve for pressure reduction. The vapor portion of the low temperature separator is fed to the expander. Consequently, two benefits accrue: (1) work is produced and (2) the stream is cooled sufficiently to be used as reflux. Our target is to maintain the same recovery (99% of methane in top and 75% of ethane in bottom) and operating variables (e.g., temperature, pressure, and purity) of the top and bottom products. By having the same recovery and operating variables, the operations of the sequential columns will not be altered. From this perspective, the bottom product temperature is -140 °F which is cooler than the required temperature. Therefore, the bottom product is used to cool the feed stream.

A steady state ASPEN Plus simulation is done for the proposed turbo-expansion activity. It gives 16,000 hp (11,900 kW) as produced power from the expander. To assess the cost/benefit of the proposed system, the fixed cost of the turbo-expander systems is estimated to calculate the payback period. The fixed cost is weighed versus the savings in refrigeration usage in the condenser, power generated, and heating utility in the reboiler.

- Savings in refrigeration:

Since the overhead temperature is -140 °F, the cost of refrigeration using ethylene will be considered.

$$\frac{133MMBtu}{h} \times \frac{\$30}{MMBtu} \times \frac{24h}{1day} \times \frac{365day}{1yr} = \frac{\$34952400}{yr}$$

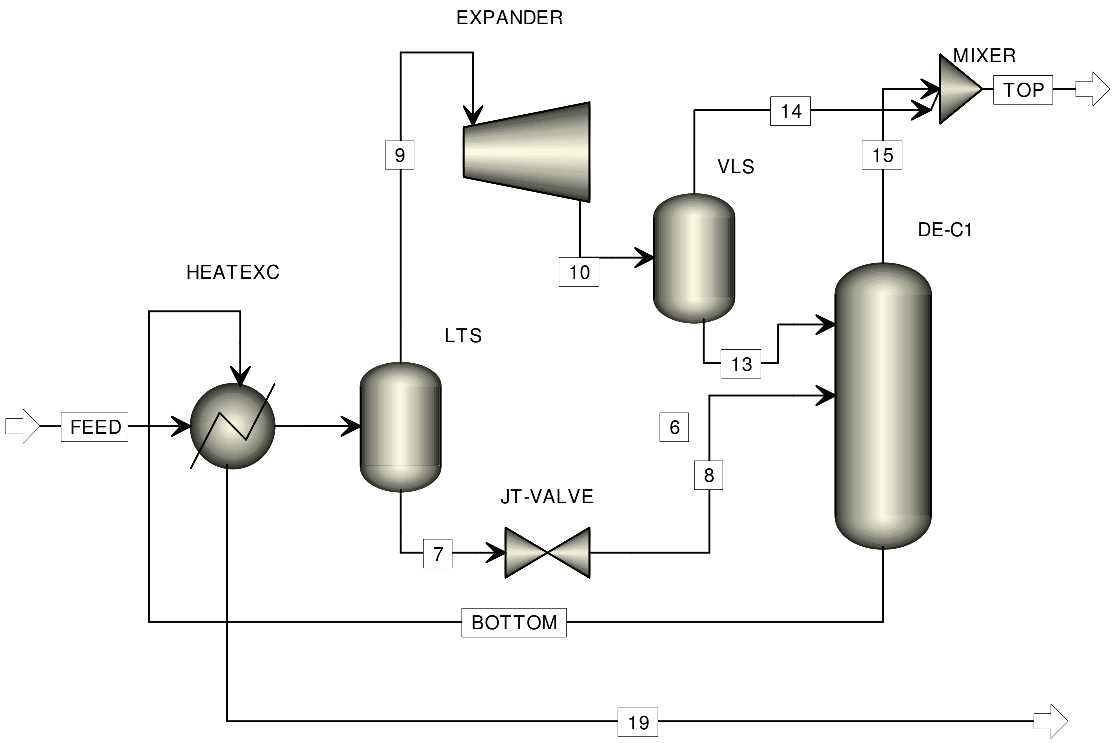

Figure 6.14 Proposed strategy for the turbo-expansion activity.

- Saving in heating utility:

$$\frac{25MMBtu}{h}\times\frac{\$6}{MMBtu}\times\frac{24h}{1day}\times\frac{365day}{1yr}=\frac{\$1314000}{yr}$$

- Power recovered:

$$11900kW\times\frac{\$0.045}{kWh}\times\frac{24h}{1day}\times\frac{365day}{1yr}=\frac{\$4690980}{yr}$$

Therefore, the total annual saving = \$ 34,952,400 + \$ 1,314,000+ \$ 4,690,980$\cong$ \$ 41×10^{6}/yr

The purchased cost is \$700,000 estimated in 2002 with a Marshall and Swift (M & S) index of 1116.9 (Peters et al. 2003).

By considering the Lang factor which is ~5 times the purchased cost, the fixed cost is \$3,500,000 (Peters et al. 2003)

Present cost = original cost (index value at present / index value of original time)

M & S index is 1410.0 for 2007 (Chemical Engineering Journal 2007).

So, the present cost = \$3,500,000 × (1410.0 / 1116.9) = \$4,418,500

$$\text{Payback period} = \frac{\$4{,}418{,}500}{\frac{\$41000000}{yr}} = 0.1yr$$

The payback period is very attractive and the expansion activity is worth applying. So, up to this stage both the retrofitting and turbo-expansion activities will be implemented as shown in Figure 6.15.

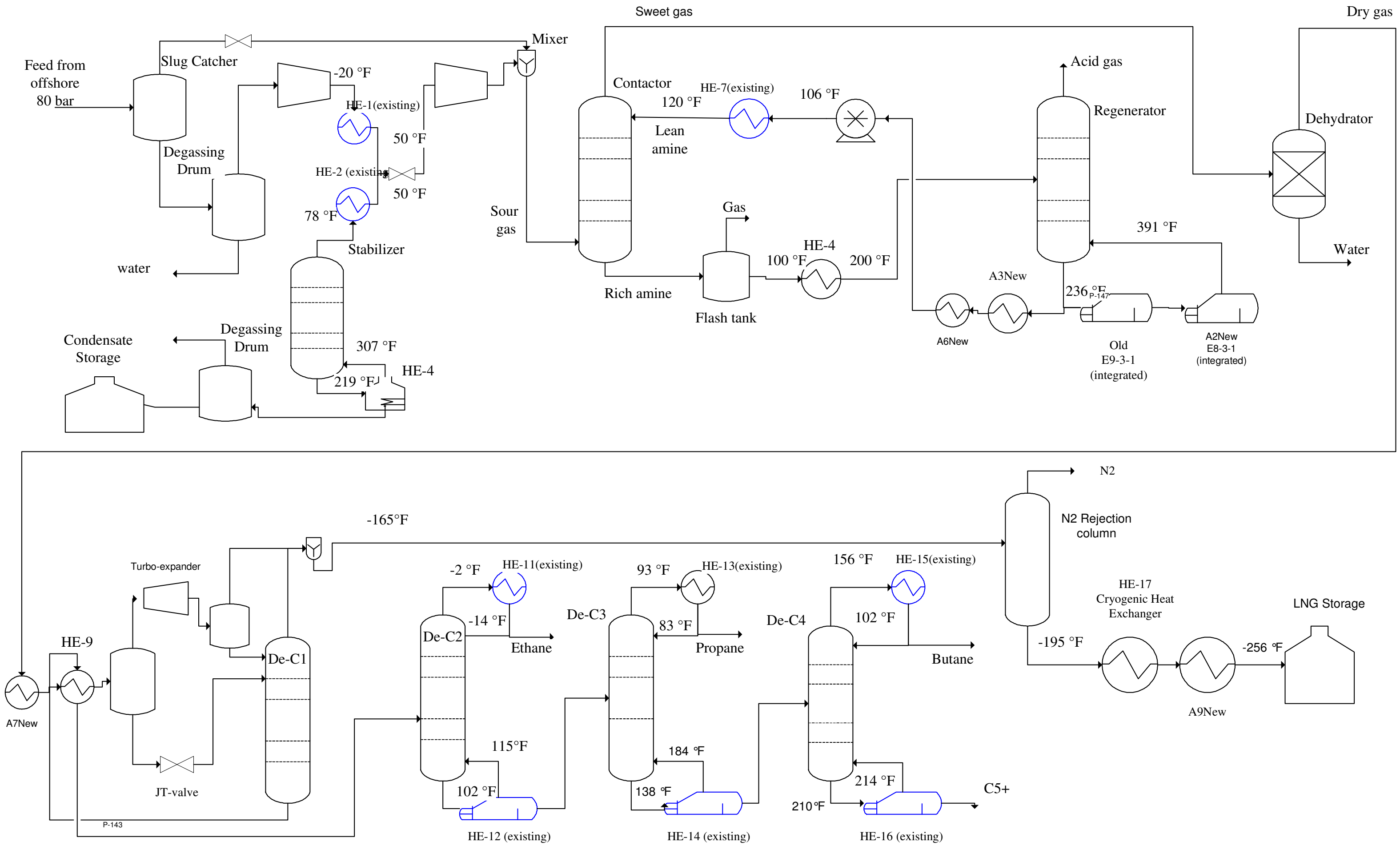

Figure 6.15 Optimized flowsheet implementing retrofitting and turbo-expansion activities.

6.7 Cogeneration Activity Results

By recalling the grand composite curve, it can be seen that two heating utilities will be sufficient to fulfill the heating demand of the process; high pressure steam and low pressure steam. Currently, the high and low pressure steams are at 235 psia and 70 psia, respectively. Accordingly, two boilers are used to supply the process with steams as shown in Figure 6.16.

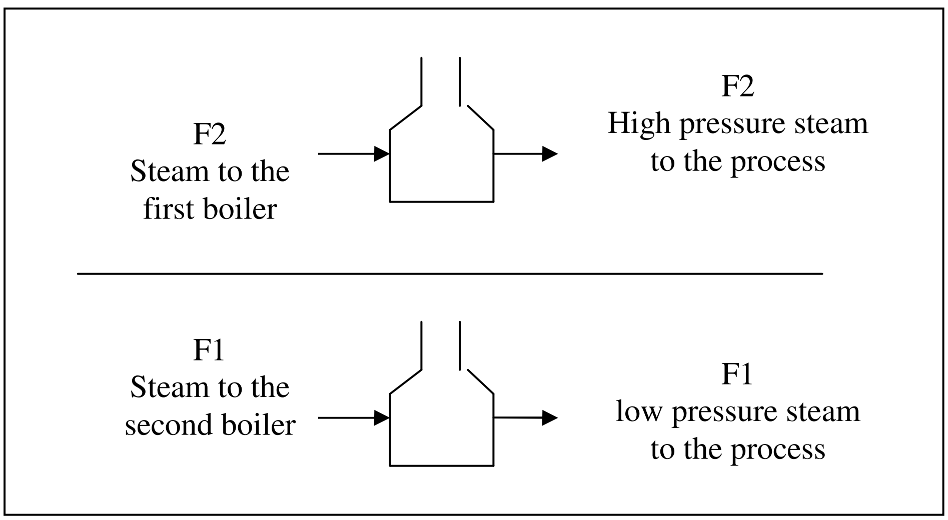

Figure 6.16 A schematic representation of process reboilers.

Now, the proposed strategy suggests feeding both steam flow rate to the first boiler. It is assumed that the first boiler is large enough and has the capacity to handle both streams. The high pressure steam is still produced as before and fed to the process, whereas the low pressure steam is produced by let the pressure down of the remaining steam flowrate through a turbine as shown in Figure 6.17.

- Steam flow rates:

$$F1 \cong \frac{Q1}{\Delta H_{lv} @ P1} = \frac{\frac{1350 MMBtu}{h}}{\frac{816.2 Btu}{Ibm}} = \frac{1654006 Ibm}{h}$$

$$F2 \cong \frac{Q2}{\Delta H_{lv} @ P2} = \frac{\frac{732.8 MMBtu}{h}}{\frac{886.9 Btu}{Ibm}} = \frac{826249 Ibm}{h}$$

By knowing the flow rates of steam, a steady state ASPEN Plus simulation is done for the proposed cogeneration activity. It gives 55,875 hp (41,666 kW) as produced power from the turbine.

The purchased cost is $500,000 estimated in 2002 with a Marshall and Swift (M & S) index of 1116.9 (Peters et al. 2003).

By considering the Lang factor which is ~5 times the purchased cost, the fixed cost is $2,500,000 (Peters et al. 2003).

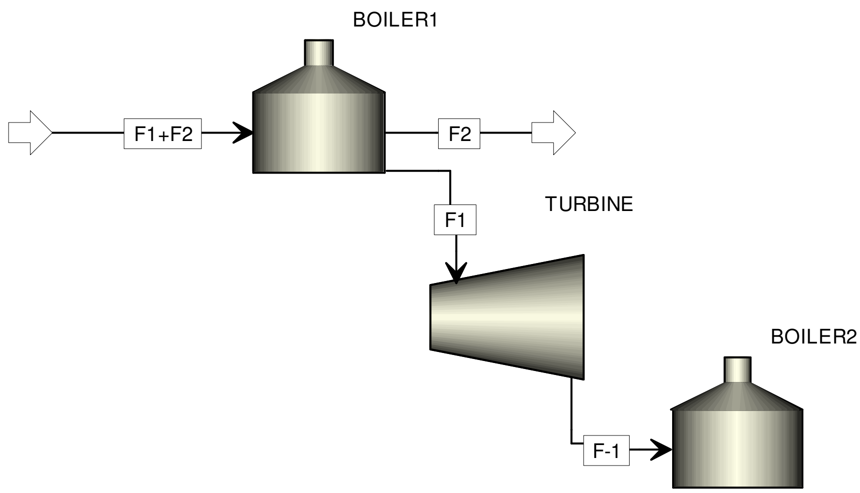

Figure 6.17 Proposed strategy for cogeneration activity.

Present cost = original cost (index value at present / index value of original time)

M & S index is 1410.0 for 2007 (Chemical Engineering Journal 2007).

So, the present cost = \$2,500,000 × (1410.0 / 1116.9) = \$3,150,000

- Power recovered:

$$41666 kW \times \frac{\$0.045}{kWh} \times \frac{24h}{1day} \times \frac{365day}{1yr} = \frac{\$16,424,700}{yr}$$

Pay back period = $\dfrac{\$3,150,000}{\dfrac{\$16,424,700}{yr}} = 0.2\,yr$

The pay back period is very attractive and suggests that the cogeneration alternative should be pursued.

7. CONCLUSIONS AND RECOMMENDATIONS

This work has provided a framework for analyzing and improving the performance of LNG plants. In particular, the following tasks have been undertaken:

- A typical LNG process flowsheet has been synthesized.
- A steady state simulation has been run using ASPEN Plus.
- The design and operating conditions for the fractionation train has been optimized by reconciling the precooling, the types and quantities of the various heating and cooling utilities, and the sizes of the columns.
- A site wide thermal pinch analysis has been applied.
- Synthesis of an HEN that satisfies the minimum heating and cooling utility targets. An MILP will be coded in LINGO and solved to identify matches between hot and cold streams along with their exchangeable heat loads.
- A new optimization formulation has been developed and applied for the retrofitting of the heat exchangers. This new approach provides a convenient framework for integrating network optimization with detailed simulation. The techniques has enabled the retrofitting of the existing heat exchangers to attain the desired heating and cooling utility targets while minimizing the installation of new heat exchangers. A combination of mathematical-programming techniques and inspection will be used to construct the retrofitted network. Also, ASPEN Plus and ICARUS were used to evaluate the performance and cost of the retrofitted heat exchangers

- ➢ Cogeneration has been examined to asses the merits of optimizing combined heat and power. The key idea is to relate the heating and cooling profiles (obtained from the thermal pinch analysis) and the steam headers to the possible introduction of steam turbines that deliver power while satisfying the heat requirements of the process.
- ➢ Turbo-expanders: The potential use of a turbo-expander system has been assessed to evaluate its impact on the refrigeration system of the process and to optimize the relationship between power and cooling utilities.

A case study on a 1,500 MMSCFD LNG plant has been carried. Simulation, optimization, and integration activities have been applied. The following are key results and observations from the case study:

- 15 % is the target for savings in the heating utility.
- 29 % is the target for savings in the cooling utility.
- Each of the three proposed energy-reduction alternatives (HEN retrofitting, turbo-expansion, and cogeneration) have very attractive payback period (less than one year), therefore suggesting that these alternatives should be pursued.

The following recommendations are suggested for future work:

- Mass integration to conserve material resources, optimize solvent usage, and manage water/wastewater
- Strategies to reduce greenhouse gas (GHG) emissions from the LNG plant
- Combined mass and heat integration to reconcile the mass and energy objectives in the process
- Flexibility analysis to examine the changes in the process design and operation with changing production rates and/or quality of the feedstocks
- Process retrofitting to address expansion needs of the process
- Integration of LNG plants with gas-to-liquid (GTL) plants

REFERENCES

Center of Liquefied Natural Gas (2007) About LNG. CLNG. http://www.lngfacts.org/About-LNG/Overview.asp. Last Updated: n/a. Access: 18 June 2007

Chemical Engineering Journal (2007) Marshall and swift equipment cost index. Chemical Engineering Journal 114(4):84

El-Halwagi MM (2006) Process integration. Elsevier, San Diego

El-Halwagi MM, Manousiouthakis V (1990) Automatic synthesis of mass exchanger networks with single-component target. Chemical Engineering Science 45(9):2813-2831

El-Halwagi MM, Manousiouthakis V (1989) Synthesis of mass exchanger networks. AIChE Journal 35(8): 1233-1244

Energy Information Administration (2007) Natural gas. EIA. http://www.eia.doe.gov/oil_gas/natural_gas/info_glance/natural_gas.html. Last Updated: n/a. Access:15 June 2007

Gas Processors Suppliers Association (1994) Engineering Data Book. Tulsa

Hamad A, Al-Fadala H, Warsame A (2007) Optimum waste interception in liquefied natural gas processes. International Journal of Environment and Pollution 29: 47-69

Hudson HM, Wilkinson JD, Cuella KT, Pierce MC (2004) Integrated liquids recovery technology improves LNG production efficiency. Spring national meeting, New Orleans

Kidnay AJ, Parrish WR (2006) Fundamentals of natural gas processing. Taylor and Francis group, Boca Raton, FL

Konukman AS, Akman U (2005) Flexibility and operability analysis of a HEN-integrated natural gas expander plant. Chemical Engineering Science 60: 7057-7074

Linnhoff B, Hindmarsh E (1983) The pinch design method for heat exchanger network. Chemical Engineering Science 38: 745-763

Linnhoff B, Flower JR (1978) Synthesis of heat exchanger networks: I. systematic generation of energy optimal networks. AIChE Journal 24: 633-642

Papoulias S, Grossmann IE (1983) A structural optimization approach in process synthesis-II: heat recovergy networks. Computers and Chemical Engineering 7(6): 707-721

Peters MS, Timmerhaus KD, West RE (2003) Plant design and economics for chemical engineers. McGraw-Hill, New York

SRI Consulting (2003) Base load LNG by cascade refrigeration. SRI Consulting. http://www.sriconsulting.com/PEP/Reports/Phase_2003/RW2003-15/RW2003-15.html. Last Updated: n/a. Access: 25 June 2007

Qatargas (2002) Feed composition of natural gas. Qatargas Operating Company Limited, Doha, State of Qatar

APPENDIX A

A.1 The Matching Formulation Code in LINGO

```
! The objective function;

min=E9_3_1+E9_1_1+E9_8_1+E9_2_1+E9_7_1+E9_4_1+E9_6_1+E8_3_1+E8_1_1
+E8_8_1 +E8_2_1 +E8_7_1 + E8_4_1 + E8_6_1+E1_3_1 +E1_1_1 +E1_8_1 +E1_2_1
+E1_7_1 + E1_4_1+E1_6_1 + E6_2_1 +E6_7_1 +E6_4_1+E6_6_1+E2_2_1 +E1_9_2
+E1_10_2+E1_5_2+E1_11_2+E1_12_2+E1_13_2+E2_9_2+E2_10_2+E2_5_2
+E2_11_2+E2_12_2+E2_13_2+E6_9_2+E6_10_2+E6_5_2+E6_11_2+E6_12_2+
E6_13_2+E5_9_2+E5_10_2+E5_5_2+E5_11_2+E5_12_2+E5_13_2+E4_11_2+
E4_12_2+E4_13_2+E3_12_2+E3_13_2+E7_13_2;


! Energy balance for hot streams;

! Energy balance for H1around temperature intervals;

R1_4 + Q1_3_4 = 70.15;
R1_5 – R1_4 + Q1_3_5 =1.15;
R1_6 – R1_5 + Q1_3_6 =19.55;
R1_7 – R1_6 + Q1_3_7 + Q1_1_7 =80.50;
R1_8 – R1_7 + Q1_1_8 =20.70;
R1_9 – R1_8 =5.75;
R1_10 – R1_9 + Q1_8_10 =4.60;
R1_11 – R1_10 =11.50;
R1_12 – R1_11 + Q1_2_12 =18.40;
R1_13 – R1_12 + Q1_2_13 +Q1_7_13 =37.95;
R1_14 – R1_13 + Q1_2_14 +Q1_7_14= 14.95;
R1_15 – R1_14 + Q1_2_15 =20.70;
R1_16 – R1_15 + Q1_2_16 +Q1_4_16=5.75;
R1_17 -R1_16 + Q1_2_17 + Q1_4_17 +Q1_6_17=10.35;
```

```
R1_18-R1_17 + Q1_2_18 +Q1_6_18=4.60;
R1_19-R1_18 +Q1_2_19=1.15;
- R1_19 + Q1_2_20=1.15;
R1_21=3.45;
R1_22 – R1_21 =2.30;
R1_23 – R1_22 =0.00;
R1_24 – R1_23 +Q1_9_24=0.00;
R1_25 – R1_24 =0.00;
R1_26 – R1_25 +Q1_10_26= 0.00;
R1_27 – R1_26 =0.00;
R1_28 – R1_27 +Q1_5_28= 0.00;
R1_29 – R1_28 =0.00;
R1_30 – R1_29 =0.00;
R1_31 – R1_30 =0.00;
R1_32 – R1_31 +Q1_11_32= 0.00;
R1_33 – R1_32=0.00;
R1_34 – R1_33=0.00;
R1_35 – R1_34 +Q1_12_35=0.00;
R1_36 – R1_35 =0.00;
R1_37 – R1_36 =0.00;
R1_38 – R1_37 =0.00;
-R1_38+Q1_13_39=0.00;

! Energy balance for H2 around temperature intervals;

Q2_2_1=1.61;
R2_21 = 4.83;
R2_22 - R2_21 =3.22;
R2_23 - R2_22 =11.27;
R2_24 - R2_23 + Q2_9_24 =16.10;
```

R2_25 - R2_24 =33.81;

R2_26 - R2_25 + Q2_10_26 =27.37;

R2_27 - R2_26= 49.91;

R2_28 - R2_27 + Q2_5_28 =24.15;

R2_29 - R2_28 =1.61;

R2_30 - R2_29 =19.32;

R2_31-R2_30 =32.20;

R2_32 – R2_31+Q2_11_32=1.61;

R2_33 - R2_32 =0.00;

R2_34- R2_33 =0.00;

R2_35 - R2_34 +Q2_12_35=0.00;

R2_36 - R2_35 =0.00;

R2_37 - R2_36 =0.00;

R2_38 - R2_37 =0.00;

- R2_38 +Q2_13_39= 0.00;

! Energy balance for H3 around temperature intervals;

R3_34 = 24.90;

R3_35 - R3_34 + Q3_12_35 =4.15;

R3_36 - R3_35 = 83.00;

R3_37 - R3_36 =0.00;

R3_38 - R3_37 = 0.00;

-R3_38 +Q3_13_39= 0.00;

! Energy balance for H4 around temperature intervals;

R4_30=73.20;

R4_31 - R4_30 = 0.00;

R4_32 - R4_31 +Q4_11_32= 0.00;

R4_33 - R4_32 =0.00;

R4_34 – R4_33=0.00;

R4_35 - R4_34 +Q4_12_35= 0.00;

R4_36 - R4_35 = 0.00;

R4_37 - R4_36 =0.00;

R4_38 - R4_37 = 0.00;

- R4_38 +Q4_13_39=0.00;

! Energy balance for H5 around temperature intervals;

R5_24+ Q5_9_24 =7.20;

R5_25 - R5_24 = 0.00;

R5_26 - R5_25 +Q5_10_26=0.00;

R5_27 - R5_26 = 0.00;

R5_28 - R5_27 + Q5_5_28 =0.00;

R5_29 - R5_28= 0.00;

R5_30 - R5_29 =0.00;

R5_31 - R5_30 = 0.00;

R5_32 - R5_31+Q5_11_32 = 0.00;

R5_33 - R5_32 =0.00;

R5_34 - R5_33 = 0.00;

R5_35 - R5_34 +Q5_12_35=0.00;

R5_36- R5_35 =0.00;

R5_37 –R5_36=0.00;

R5_38 – R5_37=0.00;

-R5_38+Q5_13_39=0.00;

! Energy balance for H6 around temperature intervals;

R6_14 + Q6_2_14 +Q6_7_14= 0.78;

R6_15 – R6_14 + Q6_2_15 =1.08;

R6_16 – R6_15 + Q6_2_16 +Q6_4_16=0.30;

R6_17 –R6_16 + Q6_2_17 + Q6_4_17 +Q6_6_17=0.54;

R6_18-R6_17 + Q6_2_18 +Q6_6_18=0.24;

R6_19-R6_18 +Q6_2_19=0.06;

- R6_19 +Q6_2_20=0.06;

R6_21=0.18;

R6_22 – R6_21 =0.00;

R6_23 – R6_22 =0.00;

R6_24 – R6_23 +Q6_9_24=0.00;

R6_25 – R6_24 =0.00;

R6_26 – R6_25 +Q6_10_26= 0.00;

R6_27 – R6_26 =0.00;

R6_28 – R6_27 +Q6_5_28= 0.00;

R6_29 – R6_28 =0.00;

R6_30 – R6_29 =0.00;

R6_31 – R6_30 =0.00;

R6_32 – R6_31 +Q6_11_32= 0.00;

R6_33 – R6_32=0.00;

R6_34 – R6_33=0.00;

R6_35 – R6_34 +Q6_12_35=0.00;

R6_36 – R6_35 =0.00;

R6_37 – R6_36 =0.00;

R6_38 – R6_37 =0.00;

-R6_38+Q6_13_39=0.00;

! Energy balance for H7 around temperature intervals;

R7_37 =488.67;

R7_38 - R7_37= 0.00;

- R7_38 + Q7_13_39 =0.00;

```
! Energy balance for H8 around temperature intervals;
R8_5 + Q8_3_5 =1355.67;
R8_6 - R8_5 + Q8_3_6 = 0.00;
R8_7 - R8_6+ Q8_1_7 + Q8_3_7 =0.00;
R8_8 - R8_7 + Q8_1_8 = 0.00;
R8_9 - R8_8 =0.00;
R8_10 - R8_9+Q8_8_10 = 0.00;
R8_11 - R8_10=0.00;
R8_12 - R8_11 + Q8_2_12 = 0.00;
R8_13 - R8_12 + Q8_2_13 + Q8_7_13 =0.00;
R8_14 - R8_13 + Q8_2_14+Q8_7_14 = 0.00;
R8_15 - R8_14 + Q8_2_15 =0.00;
R8_16 - R8_15 + Q8_2_16+Q8_4_16 =0.00;
R8_17 - R8_16 + Q8_2_17 + Q8_4_17+Q8_6_17 = 0.00;
R8_18 - R8_17 + Q8_2_18+Q8_6_18 =0.00;
R8_19- R8_18 +Q8_2_19=0.00;
-R8_19 +Q8_2_20=0.00;

! Energy balance for H9 around temperature intervals;
R9_1=727.13;
R9_2 - R9_1 =0.00;
R9_3 - R9_2 + Q9_3_3=0.00;
R9_4 - R9_3 + Q9_3_4=0.00;
R9_5 - R9_4 + Q9_3_5=0.00;
R9_6 - R9_5 + Q9_3_6=0.00;
R9_7 - R9_6 + Q9_1_7 + Q9_3_7=0.00;
R9_8 - R9_7 + Q9_1_8=0.00;
R9_9 - R9_8=0.00;
R9_10 - R9_9 + Q9_8_10=0.00;
R9_11 - R9_10=0.00;
```

R9_12 - R9_11 + Q9_2_12=0.00;

R9_13 - R9_12 + Q9_2_13 +Q9_7_13=0.00;

R9_14 - R9_13 + Q9_2_14 +Q9_7_14=0.00;

R9_15 - R9_14 + Q9_2_15=0.00;

R9_16 - R9_15 + Q9_2_16+Q9_4_16=0.00;

R9_17 - R9_16 + Q9_2_17 +Q9_4_17+Q9_6_17=0.00;

R9_18 - R9_17 + Q9_2_18 +Q9_6_18= 0.00;

R9_19 – R9_18 +Q9_2_19=0.00;

- R9_19 +Q9_2_20=0.00;

! Energy balance for cold streams;

! Energy balance for C1 around temperature intervals;
Q1_1_7+Q8_1_7+Q9_1_7=39.20;

Q1_1_8+Q8_1_8+Q9_1_8=10.08;

! Energy balance for C2 around temperature intervals;
Q1_2_12+Q8_2_12+Q9_2_12=65.28;

Q1_2_13+Q8_2_13+Q9_2_13=134.64;

Q1_2_14+Q6_2_14+Q8_2_14+Q9_2_14=53.04;

Q1_2_15+Q6_2_15+Q8_2_15+Q9_2_15=73.44;

Q1_2_16+Q6_2_16+Q8_2_16+Q9_2_16=20.40;

Q1_2_17+Q6_2_17+Q8_2_17+Q9_2_17=36.72;

Q1_2_18+Q6_2_18+Q8_2_18+Q9_2_18=16.32;

Q1_2_19+Q6_2_19+Q8_2_19+Q9_2_19=4.08;

Q1_2_20+Q2_2_20+Q6_2_20+Q8_2_20+Q9_2_20=4.08;

! Energy balance for C3around temperature intervals;
Q9_3_3=60.40;

Q9_3_4+Q1_3_4=736.88;

Q9_3_5+Q1_3_5+Q8_3_5=12.08;

Q9_3_6+Q1_3_6+Q8_3_6=205.36;

Q9_3_7+Q1_3_7+Q8_3_7=845.60;

! Energy balance for C4 around temperature intervals;
Q9_4_16+Q1_4_16+Q8_4_16 +Q6_4_16=4.95;

Q9_4_17+Q1_4_17+Q8_4_17 +Q6_4_17=8.91;

! Energy balance for C5 around temperature intervals;
Q1_5_28+Q2_5_28+Q5_5_28+ Q6_5_28=24.75;

! Energy balance for C6 around temperature intervals;
Q9_6_17+Q1_6_17+Q8_6_17 +Q6_6_17=51.93;

Q9_6_18+Q1_6_18+Q8_6_18 +Q6_6_18=23.08;

! Energy balance for C7around temperature intervals;
Q9_7_13+Q8_7_13+Q1_7_13=4.95;

Q9_7_14+Q8_7_14+Q1_7_14+Q6_7_14=1.95;

! Energy balance for C8 around temperature intervals;
Q9_8_10+Q1_8_10+Q8_8_10=3.00;

! Energy balance for C9 around temperature intervals;
Q1_9_24+Q2_9_24+Q5_9_24+Q6_9_24=4.85;

! Energy balance for C10 around temperature intervals;
Q1_10_26+Q2_10_26+Q5_10_26+Q6_10_26=6.18;

! Energy balance for C11 around temperature intervals;
Q1_11_32+Q2_11_32+Q5_11_32+Q6_11_32+Q4_11_32=177.62;

! Energy balance for C12 around temperature intervals;

Q1_12_35 + Q2_12_35 + Q3_12_35 + Q4_12_35 + Q5_12_35 +Q6_12_35= 28.68;

! Energy balance for C13 around temperature intervals;
Q1_13_39 + Q2_13_39 + Q3_13_39 + Q4_13_39 + Q5_13_39
+Q6_13_39+Q7_13_39= 571.67;

!Matching of Loads;
Q9_3_3 + Q9_3_4 + Q9_3_5 + Q9_3_6+Q9_3_7 <= 727.13*E9_3_1;
Q9_1_7 + Q9_1_8 <=49.28*E9_1_1;
Q9_8_10<=3.00*E9_8_1;
Q9_2_12+Q9_2_13+Q9_2_14+Q9_2_15+Q9_2_16+Q9_2_17+Q9_2_18+Q9_2_19+Q9
_2_20 <= 408.00*E9_2_1;
Q9_7_13 + Q9_7_14<= 6.90*E9_7_1;
Q9_4_16 + Q9_4_17 <= 13.86*E9_4_1;
Q9_6_17+Q9_6_18<=75.01*E9_6_1;
Q8_3_3 + Q8_3_4 + Q8_3_5 + Q8_3_6 +Q8_3_7 <= 1355.67*E8_3_1;
Q8_1_7 + Q8_1_8 <=49.28*E8_1_1;
Q8_8_10<=3.00*E8_8_1;
Q8_2_12+Q8_2_13+Q8_2_14+Q8_2_15+Q8_2_16+Q8_2_17+Q8_2_18+Q8_2_19+Q8
_2_20 <= 408.00*E8_2_1;
Q8_7_13 + Q8_7_14<= 6.90*E8_7_1;
Q8_4_16 + Q8_4_17 <= 13.86*E8_4_1;
Q8_6_17+Q8_6_18<=75.01*E8_6_1;
Q1_3_3 + Q1_3_4 + Q1_3_5 + Q1_3_6 +Q1_3_7 <= 334.65*E1_3_1;
Q1_1_7 + Q1_1_8 <=49.28*E1_1_1;
Q1_8_10<=3.00*E1_8_1;
Q1_2_12+Q1_2_13+Q1_2_14+Q1_2_15+Q1_2_16+Q1_2_17+Q1_2_18+Q1_2_19+Q1
_2_20 <= 334.65*E1_2_1;
Q1_7_13 + Q1_7_14<= 6.90*E1_7_1;
Q1_4_16 + Q1_4_17 <= 13.86*E1_4_1;

```
Q1_6_17+Q1_6_18<=75.01*E1_6_1;
Q6_2_14+Q6_2_15+Q6_2_16+Q6_2_17+Q6_2_18+Q6_2_19+Q6_2_20 <=
3.24*E6_2_1;
Q6_7_14<= 0.78*E6_7_1;
Q6_4_16 + Q6_4_17 <= 0.84*E6_4_1;
Q6_6_17+Q6_6_18<=3.24*E6_6_1;

Q2_2_20<=1.61*E2_2_1;
Q1_9_24<=5.75*E1_9_2;
Q1_10_26<=5.75*E1_10_2;
Q1_5_28<=5.75*E1_5_2;
Q1_11_32<=5.75*E1_11_2;
Q1_12_35<=5.75*E1_12_2;
Q1_13_39<=5.75*E1_13_2;
Q2_9_24<=48.55*E2_9_2;
Q2_10_26<=61.18*E2_10_2;
Q2_5_28<=24.75*E2_5_2;
Q2_11_32<=177.62*E2_11_2;
Q2_12_35<=28.68*E2_12_2;
Q2_13_39<=225.40*E2_13_2;
Q6_9_24<=0.18*E6_9_2;
Q6_10_26<=0.18*E6_10_2;
Q6_5_28<=0.18*E6_5_2;
Q6_11_32<=0.18*E6_11_2;
Q6_12_35<=0.18*E6_12_2;
Q6_13_39<=0.18*E6_13_2;
Q5_9_24<=7.20*E5_9_2;
Q5_10_26<=7.20*E5_10_2;
Q5_5_28<=7.20*E5_5_2;
```

```
Q5_11_32<=7.20*E5_11_2;
Q5_12_35<=7.20*E5_12_2;
Q5_13_39<=7.20*E5_13_2;
Q4_11_32<=73.20*E4_11_2;
Q4_12_35<=28.68*E4_12_2;
Q4_13_39<=73.20*E4_13_2;
Q3_12_35<=28.68*E3_12_2;
Q3_13_39<=112.05*E3_13_2;
Q7_13_39<=488.67*E7_13_2;
! Non-negative residuals;
R1_4 > 0.00;
R1_5 > 0.00;
R1_6 > 0.00;
R1_7 > 0.00;
R1_8 > 0.00;
R1_9> 0.00;
R1_10 > 0.00;
R1_11> 0.00;
R1_12 > 0.00;
R1_13 > 0.00;
R1_14 > 0.00;
R1_15 > 0.00;
R1_16 > 0.00;
R1_17 > 0.00;
R1_18>0.00;
R1_19 > 0.00;
R1_21 > 0.00;
R1_22> 0.00;
R1_23> 0.00;
R1_24> 0.00;
```

R1_25 > 0.00;

R1_26 > 0.00;

R1_27 > 0.00;

R1_28> 0.00;

R1_29 > 0.00;

R1_30 > 0.00;

R1_31 > 0.00;

R1_32 > 0.00;

R1_33 > 0.00;

R1_34 > 0.00;

R1_35> 0.00;

R1_36 > 0.00;

R1_37 > 0.00;

R1_38 > 0.00;

R2_21 > 0.00;

R2_22> 0.00;

R2_23> 0.00;

R2_24> 0.00;

R2_25 > 0.00;

R2_26 > 0.00;

R2_27 > 0.00;

R2_28> 0.00;

R2_29 > 0.00;

R2_30 > 0.00;

R2_31 > 0.00;

R2_32 > 0.00;

R2_33 > 0.00;

R2_34 > 0.00;

R2_35 > 0.00;

R2_36 > 0.00;

R2_37 > 0.00;

R2_38 > 0.00;

R3_34 > 0.00;

R3_35>0.00;

R3_36 > 0.00;

R3_37>0.00;

R3_38 > 0.00;

R4_30 > 0.00;

R4_31 > 0.00;

R4_32 > 0.00;

R4_33> 0.00;

R4_34 > 0.00;

R4_35 > 0.00;

R4_36 > 0.00;

R4_37 > 0.00;

R4_38 > 0.00;

R5_24 >0.00;

R5_25> 0.00;

R5_26 >0.00;

R5_27> 0.00;

R5_28 >0.00;

R5_29> 0.00;

R5_30 >0.00;

R5_31> 0.00;

R5_32 >0.00;

R5_33 >0.00;

R5_34> 0.00;

R5_35 >0.00;

R5_36 >0.00;

R5_37> 0.00;

R5_38 >0.00;

R6_14 > 0.00;

R6_15 > 0.00;

R6_16 > 0.00;

R6_17 > 0.00;

R6_18>0.00;

R6_19 > 0.00;

R6_21 > 0.00;

R6_22> 0.00;

R6_23> 0.00;

R6_24> 0.00;

R6_25 > 0.00;

R6_26 > 0.00;

R6_27 > 0.00;

R6_28> 0.00;

R6_29 > 0.00;

R6_30 > 0.00;

R6_31 > 0.00;

R6_32 > 0.00;

R6_33 > 0.00;

R6_34 > 0.00;

R6_35> 0.00;

R6_36 > 0.00;

R6_37 > 0.00;

R6_38 > 0.00;

R7_37 > 0.00;

R7_38 > 0.00;

R8_5 > 0.00;

R8_6 > 0.00;

R8_7 > 0.00;

R8_8 > 0.00;

R8_9 > 0.00;

R8_10 > 0.00;

R8_11 > 0.00;

R8_12 > 0.00;

R8_13> 0.00;

R8_14 > 0.00;

R8_15 > 0.00;

R8_16 > 0.00;

R8_17 > 0.00;

R8_18 >0.00;

R8_19 > 0.00;

R9_1 > 0.00;

R9_2 > 0.00;

R9_3 > 0.00;

R9_4> 0.00;

R9_5 > 0.00;

R9_6 > 0.00;

R9_7 > 0.00;

R9_8 > 0.00;

R9_9 > 0.00;

R9_10 > 0.00;

R9_11 > 0.00;

R9_12 > 0.00;

R9_13 > 0.00;

R9_14 > 0.00;

R9_15 > 0.00;

R9_16 > 0.00;

R9_17 > 0.00;

R9_18 > 0.00;

R9_19 >0.00;

! Non-negative loads;
Q1_1_7>0.00;

Q8_1_7>0.00;

Q9_1_7>0.00;

Q1_1_8>0.00;

Q8_1_8>0.00;

Q9_1_8>0.00;

Q1_2_12>0.00;

Q8_2_12>0.00;

Q9_2_12>0.00;

Q1_2_13>0.00;

Q8_2_13>0.00;

Q9_2_13>0.00;

Q1_2_14>0.00;

Q6_2_14>0.00;

Q8_2_14>0.00;

Q9_2_14>0.00;

Q1_2_15>0.00;

Q6_2_15>0.00;

Q8_2_15>0.00;

Q9_2_15>0.00;

Q1_2_16>0.00;

Q6_2_16>0.00;

Q8_2_16>0.00;

Q9_2_16>0.00;

Q1_2_17>0.00;

Q6_2_17>0.00;

Q8_2_17>0.00;

Q9_2_17>0.00;

Q1_2_18>0.00;

Q6_2_18>0.00;

Q8_2_18>0.00;

Q9_2_18>0.00;

Q1_2_19>0.00;

Q6_2_19>0.00;

Q8_2_19>0.00;

Q9_2_19>0.00;

Q1_2_20>0.00;

Q2_2_20>0.00;

Q6_2_20>0.00;

Q8_2_20>0.00;

Q9_2_20>0.00;

Q9_3_3>0.00;

Q9_3_4>0.00;

Q1_3_4>0.00;

Q9_3_5>0.00;

Q1_3_5>0.00;

Q8_3_5>0.00;

Q9_3_6>0.00;

Q1_3_6>0.00;

Q8_3_6>0.00;

Q9_3_7>0.00;

Q1_3_7>0.00;

Q8_3_7>0.00;

Q9_4_16>0.00;

Q1_4_16>0.00;

Q8_4_16>0.00;

Q6_4_16>0.00;

Q9_4_17>0.00;

Q1_4_17>0.00;

Q8_4_17>0.00;

Q6_4_17>0.00;

Q1_5_28>0.00;

Q2_5_28>0.00;

Q5_5_28>0.00;

Q6_5_28>0.00;

Q9_6_17>0.00;

Q1_6_17>0.00;

Q8_6_17>0.00;

Q6_6_17>0.00;

Q9_6_18>0.00;

Q1_6_18>0.00;

Q8_6_18>0.00;

Q6_6_18>0.00;

Q9_7_13>0.00;

Q8_7_13>0.00;

Q1_7_13>0.00;

Q9_7_14>0.00;

Q8_7_14>0.00;

Q1_7_14>0.00;

Q6_7_14>0.00;

Q9_8_10>0.00;

```
Q1_8_10>0.00;
Q8_8_10>0.00;
Q1_9_24>0.00;
Q2_9_24>0.00;
Q5_9_24>0.00;
Q6_9_24>0.00;
Q1_10_26>0.00;
Q2_10_26>0.00;
Q5_10_26>0.00;
Q6_10_26>0.00;
Q1_11_32>0.00;
Q2_11_32>0.00;
Q5_11_32>0.00;
Q6_11_32>0.00;
Q4_11_32>0.00;
Q1_12_35>0.00;
Q2_12_35>0.00;
Q3_12_35>0.00;
Q4_12_35>0.00;
Q5_12_35>0.00;
Q6_12_35>0.00;
Q1_13_39>0.00;
Q2_13_39>0.00;
Q3_13_39>0.00;
Q4_13_39>0.00;
Q5_13_39>0.00;
Q6_13_39>0.00;
Q7_13_39>0.00;
```

```
! Binary integer variables for matching streams;
@bin (E9_3_1);
@bin(E9_1_1);
@bin (E9_8_1);
@bin (E9_2_1);
@bin (E9_7_1);
@bin (E9_4_1);
@bin (E9_6_1);
@bin (E8_3_1);
@bin (E8_1_1);
@bin (E8_8_1);
@bin (E8_2_1);
@bin (E8_7_1);
@bin (E8_4_1);
@bin (E8_6_1);
@bin (E1_3_1);
@bin (E1_1_1);
@bin (E1_8_1);
@bin (E1_2_1);
@bin (E1_7_1);
@bin (E1_4_1);
@bin (E1_6_1);
@bin (E6_2_1);
@bin (E6_7_1);
@bin (E6_4_1);
@bin (E6_6_1);
@bin (E2_2_1);
@bin (E1_9_2);
@bin (E1_10_2);
@bin (E1_5_2);
```

```
@bin (E1_11_2);
@bin (E1_12_2);
@bin (E1_13_2);
@bin (E2_9_2);
@bin (E2_10_2);
@bin (E2_5_2);
@bin (E2_11_2);
@bin (E2_12_2);
@bin (E2_13_2);
@bin (E6_9_2);
@bin (E6_10_2);
@bin (E6_5_2);
@bin (E6_11_2);
@bin (E6_12_2);
@bin (E6_13_2);
@bin (E5_9_2);
@bin (E5_10_2);
@bin (E5_5_2);
@bin (E5_11_2);
@bin (E5_12_2);
@bin (E5_13_2);
@bin (E4_11_2);
@bin (E4_12_2);
@bin (E4_13_2);
@bin (E3_12_2);
@bin (E3_13_2);
@bin (E7_13_2);
END
```

A.2 The Matching Formulation Solution from LINGO

```
Global optimal solution found.
  Objective value:                              16.00000
  Extended solver steps:                               0
  Total solver iterations:                            46
```

Variable	Value	Reduced Cost
E9_3_1	1.000000	0.000000
E9_7_2	0.000000	1.000000
E9_4_1	0.000000	1.000000
E9_6_1	0.000000	1.000000
E8_3_1	1.000000	1.000000
E8_1_1	1.000000	1.000000
E8_8_1	1.000000	1.000000
E8_2_1	1.000000	1.000000
E8_4_1	1.000000	1.000000
E8_7_1	0.000000	1.000000
E8_6_1	1.000000	1.000000
E1_3_1	1.000000	1.000000
E1_8_1	0.000000	1.000000
E1_1_1	0.000000	1.000000
E1_2_1	1.000000	1.000000
E1_7_1	1.000000	1.000000
E1_4_1	0.000000	1.000000
E1_6_1	0.000000	1.000000
E6_6_1	0.000000	1.000000
E1_9_2	0.000000	1.000000
E1_10_2	0.000000	1.000000
E2_9_2	0.000000	1.000000
E2_10_2	1.000000	1.000000
E2_5_2	1.000000	1.000000
E2_11_2	1.000000	1.000000
E6_9_2	0.000000	1.000000
E6_10_2	0.000000	1.000000
E5_9_2	1.000000	1.000000
E5_10_2	0.000000	1.000000
E4_11_2	0.000000	1.000000
E3_12_2	0.000000	1.000000
E3_13_2	1.000000	1.000000
E7_13_2	1.000000	1.000000
R1_4	0.000000	0.000000
Q1_3_4	70.15000	0.000000
R1_5	0.000000	0.000000
Q1_3_5	1.150000	0.000000
R1_6	0.000000	0.000000
Q1_3_6	9.550000	0.000000
R1_7	0.000000	0.000000
Q1_3_7	0.000000	0.000000
Q1_1_7	0.000000	0.000000
R1_8	20.70000	0.000000

Q1_1_8	0.000000	0.000000
R1_9	26.45000	0.000000
R1_10	31.05000	0.000000
Q1_8_10	0.000000	0.000000
R1_11	42.55000	0.000000
R1_12	0.000000	0.000000
Q1_2_12	60.95000	0.000000
R1_13	33.00000	0.000000
Q1_2_13	0.000000	0.000000
Q1_7_13	4.950000	0.000000
R1_14	46.78000	0.000000
Q1_2_14	0.000000	0.000000
Q1_7_14	1.170000	0.000000
R1_15	16.02000	0.000000
Q1_2_15	51.46000	0.000000
R1_16	1.370000	0.000000
Q1_2_16	20.40000	0.000000
Q1_4_16	0.000000	0.000000
R1_17	11.72000	0.000000
Q1_2_17	0.000000	0.000000
Q1_4_17	0.000000	0.000000
Q1_6_17	0.000000	0.000000
R1_18	0.000000	0.000000
Q1_2_18	16.32000	0.000000
Q1_6_18	0.000000	0.000000
R1_19	0.000000	0.000000
Q1_2_19	1.150000	0.000000
Q1_2_20	1.150000	0.000000
R1_21	3.450000	0.000000
R1_22	5.750000	0.000000
R1_23	5.750000	0.000000
R1_24	5.750000	0.000000
Q1_9_24	0.000000	0.000000
R1_25	5.750000	0.000000
R1_26	5.750000	0.000000
Q1_10_26	0.000000	0.000000
R1_27	5.750000	0.000000
R1_28	5.750000	0.000000
Q1_5_28	0.000000	0.000000
R1_29	5.750000	0.000000
R1_30	5.750000	0.000000
R1_31	5.750000	0.000000
R1_32	0.000000	0.000000
Q1_11_32	5.750000	0.000000
R1_33	0.000000	0.000000
R1_34	0.000000	0.000000
R1_35	0.000000	0.000000
Q1_12_35	0.000000	0.000000
R1_36	0.000000	0.000000
R1_37	0.000000	0.000000
R1_38	0.000000	0.000000
Q1_13_39	0.000000	0.000000
Q2_2_1	1.610000	0.000000
R2_21	4.830000	0.000000

R2_22	8.050000	0.000000
R2_23	19.32000	0.000000
R2_24	35.42000	0.000000
Q2_9_24	0.000000	0.000000
R2_25	69.23000	0.000000
R2_26	90.42000	0.000000
Q2_10_26	6.180000	0.000000
R2_27	140.3300	0.000000
R2_28	139.7300	0.000000
Q2_5_28	24.75000	0.000000
R2_29	141.3400	0.000000
R2_30	160.6600	0.000000
R2_31	192.8600	0.000000
R2_32	25.13000	0.000000
Q2_11_32	169.3400	0.000000
R2_33	25.13000	0.000000
R2_34	25.13000	0.000000
R2_35	0.000000	0.000000
Q2_12_35	25.13000	0.000000
R2_36	0.000000	0.000000
R2_37	0.000000	0.000000
R2_38	0.000000	0.000000
Q2_13_39	0.000000	0.000000
R3_34	24.90000	0.000000
R3_35	29.05000	0.000000
Q3_12_35	0.000000	0.000000
R3_36	112.0500	0.000000
R3_37	112.0500	0.000000
R3_38	112.0500	0.000000
R4_30	73.20000	0.000000
R4_31	73.20000	0.000000
R4_32	73.20000	0.000000
Q4_11_32	0.000000	0.000000
R4_33	73.20000	0.000000
R4_34	73.20000	0.000000
R4_35	69.65000	0.000000
Q4_12_35	3.550000	0.000000
R4_36	69.65000	0.000000
R4_37	69.65000	0.000000
R4_38	69.65000	0.000000
Q4_13_39	69.65000	0.000000
R5_24	2.350000	0.000000
Q5_9_24	4.850000	0.000000
R5_25	2.350000	0.000000
R5_26	2.350000	0.000000
Q5_10_26	0.000000	0.000000
R5_27	2.350000	0.000000
R5_28	2.350000	0.000000
Q5_5_28	0.000000	0.000000
R5_29	2.350000	0.000000
R5_30	2.350000	0.000000
R5_31	2.350000	0.000000
R5_32	0.000000	0.000000
Q5_11_32	2.350000	0.000000

R5_33	0.000000	0.000000
R5_34	0.000000	0.000000
R5_35	0.000000	0.000000
Q5_12_35	0.000000	0.000000
R5_36	0.000000	0.000000
R5_37	0.000000	0.000000
R5_38	0.000000	0.000000
Q5_13_39	0.000000	0.000000
R6_14	0.000000	0.000000
Q6_2_14	0.000000	0.000000
Q6_7_14	0.7800000	0.000000
R6_15	1.080000	0.000000
Q6_2_15	0.000000	0.000000
R6_16	0.5400000	0.000000
Q6_2_16	0.000000	0.000000
Q6_4_16	0.8400000	0.000000
R6_17	1.080000	0.000000
Q6_2_17	0.000000	0.000000
Q6_4_17	0.000000	0.000000
Q6_6_17	0.000000	0.000000
R6_18	1.320000	0.000000
Q6_2_18	0.000000	0.000000
Q6_6_18	0.000000	0.000000
R6_19	0.000000	0.000000
Q6_2_19	1.380000	0.000000
Q6_2_20	0.6000000	0.000000
R6_21	0.1800000	0.000000
R6_22	0.1800000	0.000000
R6_23	0.1800000	0.000000
R6_24	0.1800000	0.000000
Q6_9_24	0.000000	0.000000
R6_25	0.1800000	0.000000
R6_26	0.1800000	0.000000
Q6_10_26	0.000000	0.000000
R6_27	0.1800000	0.000000
R6_28	0.1800000	0.000000
Q6_5_28	0.000000	0.000000
R6_29	0.1800000	0.000000
R6_30	0.1800000	0.000000
R6_31	0.1800000	0.000000
R6_32	0.000000	0.000000
Q6_11_32	0.1800000	0.000000
R6_33	0.000000	0.000000
R6_34	0.000000	0.000000
R6_35	0.000000	0.000000
Q6_12_35	0.000000	0.000000
R6_36	0.000000	0.000000
R6_37	0.000000	0.000000
R6_38	0.000000	0.000000
Q6_13_39	0.000000	0.000000
R7_37	488.6700	0.000000
R7_38	488.6700	0.000000
Q7_13_39	488.6700	0.000000
R8_5	1344.740	0.000000

Q8_3_5	10.93000	0.000000
R8_6	1158.930	0.000000
Q8_3_6	185.8100	0.000000
R8_7	354.6300	0.000000
Q8_1_7	39.20000	0.000000
Q8_3_7	765.1000	0.000000
R8_8	344.5500	0.000000
Q8_1_8	10.08000	0.000000
R8_9	344.5500	0.000000
R8_10	341.5500	0.000000
Q8_8_10	3.000000	0.000000
R8_11	341.5500	0.000000
R8_12	337.2200	0.000000
Q8_2_12	4.330000	0.000000
R8_13	202.5800	0.000000
Q8_2_13	134.6400	0.000000
Q8_7_13	0.000000	0.000000
R8_14	149.5400	0.000000
Q8_2_14	53.04000	0.000000
Q8_7_14	0.000000	0.000000
R8_15	127.5600	0.000000
Q8_2_15	21.98000	0.000000
R8_16	123.4500	0.000000
Q8_2_16	0.000000	0.000000
Q8_4_16	4.110000	0.000000
R8_17	25.89000	0.000000
Q8_2_17	36.72000	0.000000
Q8_4_17	8.910000	0.000000
Q8_6_17	51.93000	0.000000
R8_18	2.810000	0.000000
Q8_2_18	0.000000	0.000000
Q8_6_18	23.08000	0.000000
R8_19	1.260000	0.000000
Q8_2_19	1.550000	0.000000
Q8_2_20	1.260000	0.000000
R9_1	727.1300	0.000000
R9_2	727.1300	0.000000
R9_3	666.7300	0.000000
Q9_3_3	60.40000	0.000000
R9_4	0.000000	0.000000
Q9_3_4	666.7300	0.000000
R9_5	0.000000	0.000000
Q9_3_5	0.000000	0.000000
R9_6	0.000000	0.000000
Q9_3_6	0.000000	0.000000
R9_7	0.000000	0.000000
Q9_1_7	0.000000	0.000000
Q9_3_7	0.000000	0.000000
R9_8	0.000000	0.000000
Q9_1_8	0.000000	0.000000
R9_9	0.000000	0.000000
R9_10	0.000000	0.000000
Q9_8_10	0.000000	0.000000
R9_11	0.000000	0.000000

R9_12	0.000000	0.000000
Q9_2_12	0.000000	0.000000
R9_13	0.000000	0.000000
Q9_2_13	0.000000	0.000000
Q9_7_13	0.000000	0.000000
R9_14	0.000000	0.000000
Q9_2_14	0.000000	0.000000
Q9_7_14	0.000000	0.000000
R9_15	0.000000	0.000000
Q9_2_15	0.000000	0.000000
R9_16	0.000000	0.000000
Q9_2_16	0.000000	0.000000
Q9_4_16	0.000000	0.000000
R9_17	0.000000	0.000000
Q9_2_17	0.000000	0.000000
Q9_4_17	0.000000	0.000000
Q9_6_17	0.000000	0.000000
R9_18	0.000000	0.000000
Q9_2_18	0.000000	0.000000
Q9_6_18	0.000000	0.000000
R9_19	0.000000	0.000000
Q9_2_19	0.000000	0.000000
Q9_2_20	0.000000	0.000000
Q2_2_20	1.610000	0.000000
Q3_13_39	13.35000	0.000000
E9_1_1	1.000000	0.000000
E9_8_1	1.000000	0.000000
E9_2_1	1.000000	0.000000
E9_7_1	1.000000	0.000000
Q8_3_3	0.000000	0.000000
Q8_3_4	0.000000	0.000000
Q1_3_3	0.000000	0.000000
E6_2_1	1.000000	0.000000
E6_7_1	1.000000	0.000000
E6_4_1	1.000000	0.000000
E2_2_1	1.000000	0.000000
E1_5_2	1.000000	0.000000
E1_11_2	1.000000	0.000000
E1_12_2	1.000000	0.000000
E1_13_2	1.000000	0.000000
E2_12_2	1.000000	0.000000
E2_13_2	1.000000	0.000000
E6_5_2	1.000000	0.000000
E6_11_2	1.000000	0.000000
E6_12_2	1.000000	0.000000
E6_13_2	1.000000	0.000000
E5_5_2	1.000000	0.000000
E5_11_2	1.000000	0.000000
E5_12_2	1.000000	0.000000
E5_13_2	1.000000	0.000000
E4_12_2	1.000000	0.000000
E4_13_2	1.000000	0.000000

APPENDIX B

B.1 Integrated Heat Exchangers Calculations

Here, the matched results are: E9_3_1, E8_3_1, E8_1_1, E8_8_1, E8_2_1, E8_4_1, E8_6_1, E1_3_1, E1_2_1, E1_7_1, E2_10_2, E2_5_2, E2_11_2, E5_9_2, E3_13_2, and E7_13_2. As can be seen, each heat exchanger has three indexes. The first and second indexes represent the hot and cold stream number, respectively. The stream numbers are the same as those shown in temperature interval diagram (Table 6.1). The third index is given number one and two if it is above or below the pinch, respectively. Here, the inlet and outlet temperatures of the obtained heat exchangers from LINGO are calculated by using either equation 2.2 or 2.3.

For E9_3_1: The obtained loads from LINGO are:

Q9_3_3 = 60.40 MMBtu / h

Q9_3_4 = 666.73 MMBtu / h

So, heat transferred from H9 to C3 = 60.40+666.73 = 727.13 MMBtu / h

Heat balance:

$727.13 = 12.08 * (391 - t_{in})$

Solving for T3in = 331 °F

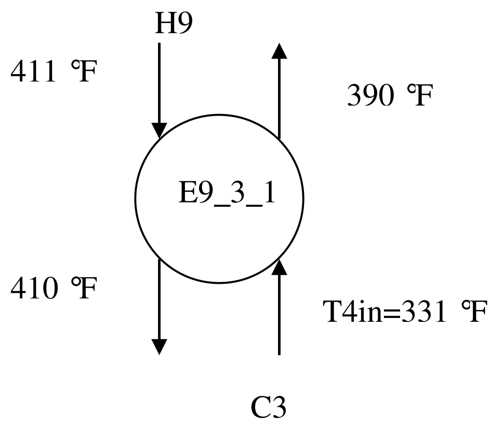

Figure B.1 Exchanger E9_3_1 calculation

For E1_3_1: The obtained loads from LINGO are:

Q1_3_4 = 70.15 MMBtu / h

Q1_3_5 = 1.15 MMBtu / h

Q1_3_6 = 9.55 MMBtu / h

So, heat transferred from H1 to C3 = 70.15+1.15+9.55 = 80.50 MMBtu / h

Heat balance:

$80.50 = 12.08 * (391 - t1_{out})$

Solving for T1out = 321 °F

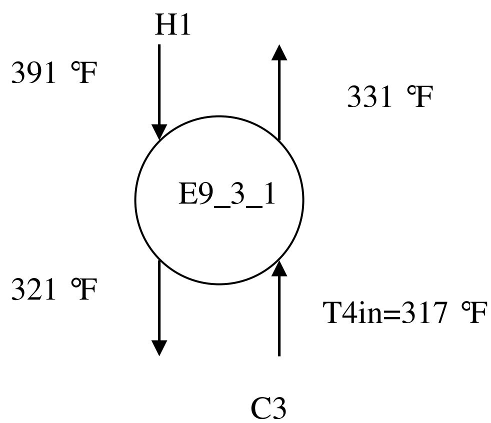

Figure B.2 Exchanger E1_3_1 calculation

For E8_3_1: The obtained loads from LINGO are:

Q8_3_5 = 10.93 MMBtu / h

Q8_3_6 = 185.81 MMBtu / h

Q8_3_7 = 765.10 MMBtu / h

So, heat transferred from H8 to C3 = 10.93+185.81+765.10 = 961.84 MMBtu / h

Heat balance:

79.6 = 12.08 * (t1out – 237)

Solving for T1out = 317 ℉

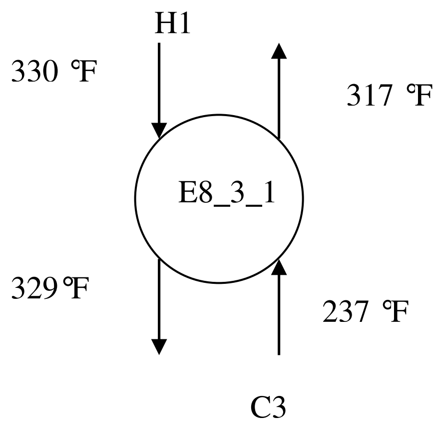

Figure B.3 Exchanger E8_3_1 calculation

For E8_1_1: The obtained loads from LINGO are:

Q8_1_7= 39.20 MMBtu / h

Q8_1_8 =10.08 MMBtu / h

So, heat transferred from H8 to C1 =39.20+10.08=49.28 MMBtu / h

Heat balance:

$49.28 = 0.56 * (307 - T1in)$

Solving for T1in = 219 ℉

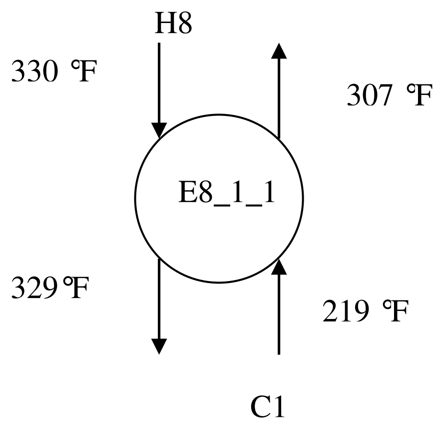

Figure B.4 Exchanger E8_1_1 calculation

For E8_8_1: The obtained loads from LINGO are:

Q8_8_10 = 3.00 MMBtu / h

So, heat transferred from H8 to C8 = 3.00 MMBtu / h

Heat balance:

3.00 = 0.75 * (t8out – 210)

Solving for T8out = 214 ℉

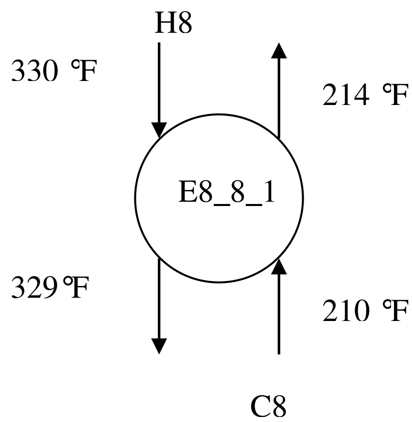

Figure B.5 Exchanger E8_8_1 calculation

For E8_4_1: The obtained loads from LINGO are:

Q8_4_16 = 4.11 MMBtu / h

Q8_4_17 = 8.91 MMBtu / h

So, heat transferred from H8 to C4= 4.11+8.91=13.02 MMBtu/h

Heat balance:

13.02 = 0.99 * (120 – t4in)

Solving for t4in = 106 ℉

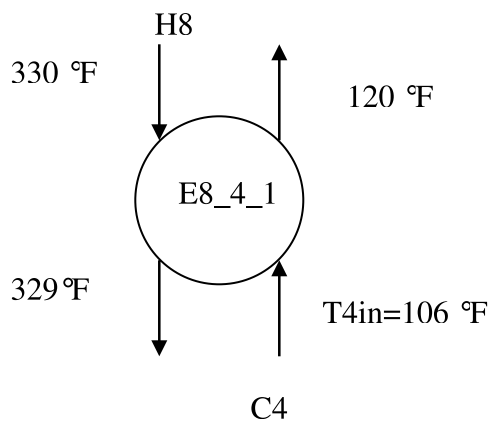

Figure B.6 Exchanger E8_4_1 calculation

For E1_7_1: The obtained loads from LINGO are:

Q1_7_13 = 4.95 MMBtu / h

Q1_7_14 = 1.17 MMBtu / h

So, heat transferred from H1 to C7= 4.95+1.17=6.12 MMBtu/h

Heat balance:

6.12 =1.15 * (248 – t1out)

Solving for t1out = 243 ℉

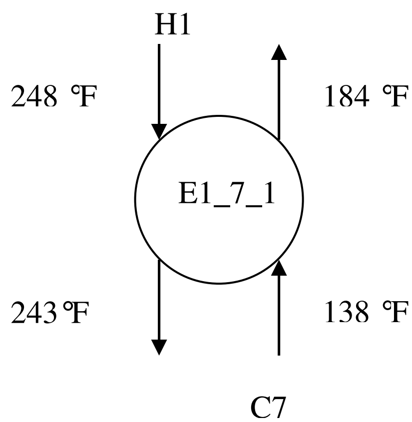

Figure B.7 Exchanger E1_7_1 calculation

For E8_6_1: The obtained loads from LINGO are:

Q8_6_17 = 51.93 MMBtu / h

Q8_6_18 = 23.08 MMBtu / h

So, heat transferred from H8 to C6= 51.93+23.08=75.01 MMBtu/h

Heat balance:

75.01 = 5.77 * (115 – t6in)

Solving for t6in = 102 ℉

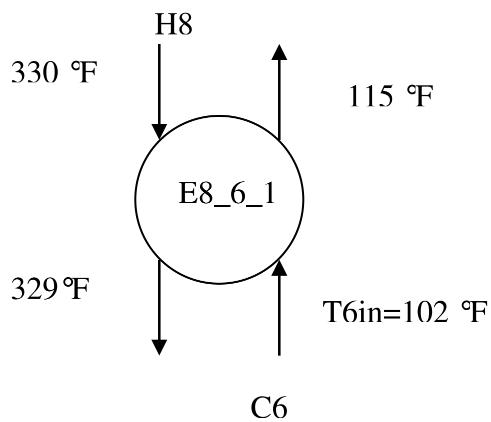

Figure B.8 Exchanger E8_6_1 calculation

Next, we move to the heat exchanges between H1 and C2 & H8 and C2 by splitting the specific heat capacity of C2:

For E1_2_1: The obtained loads from LINGO are:

Q1_2_12 = 60.95 MMBtu / h

Q1_2_15 = 51.46 MMBtu / h

Q1_2_16 = 20.40 MMBtu / h

So, heat transferred from H1 to C2 = 132.81 MMBtu / h

Heat balance:

132.81 = 1.15* (243 – T1out)

Solving for T1out = 127 ℉

For E8_2_1: The obtained loads from LINGO are:

Q8_2_12 = 4.33 MMBtu / h

Q8_2_13 = 134.04 MMBtu / h

Q8_2_14 = 53.04 MMBtu / h

Q8_2_15 = 21.98 MMBtu / h

Q8_2_17 = 36.72 MMBtu / h

So, heat transferred from H8 to C2= 250.11 MMBtu / h

Heat balance:

250.11 = 4.08 * (t2out-100)

Solving for t2out = 200 ℉

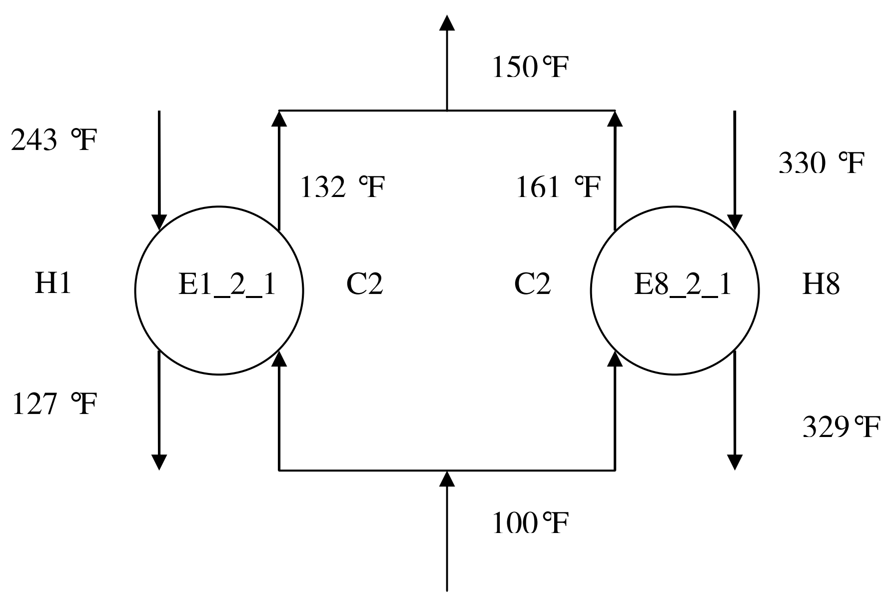

Figure B.9 Parallel Exchanging between E1_2_1 and E8_2_1calculation

Now, we will move to the heat exchangers below the pinch: E2_10_2, E2_5_2, E2_11_2, E5_9_2, E3_13_2, and E7_13_2.

For E2_10_2: The obtained loads from LINGO also are:

Q2_10_26 = 6.18 MMBtu / h

So, heat transferred from H2 to C10= 6.18 MMBtu / h

Heat balance:

6.18= 1.61* (110-T2out)

Solving for T2out = 101 ℉

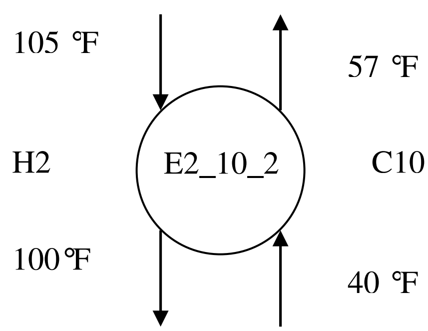

Figure B.10 Exchanger E2_10_2 calculation

For E2_5_2: The obtained loads from LINGO are:

Q2_5_21 = 24.75 MMBtu / h

So, heat transferred from H2 to C5= 24.75 MMBtu / h

Heat balance:

24.75= 1.61* (101-T2out)

Solving for T2out = 86 °F

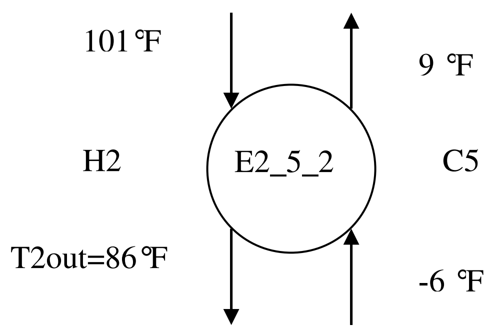

Figure B.11 Exchanger E2_5_2 calculation

For E2_11_2: The obtained loads from LINGO also are:

Q2_11_32 = 169.34 MMBtu / h

Q2_11_35 = 25.13 MMBtu / h

So, heat transferred from H2 to C11= 194.47 MMBtu / h

Heat balance:

11.7= 0.18* (86-T2out)

Solving for T2out = -35 ℉

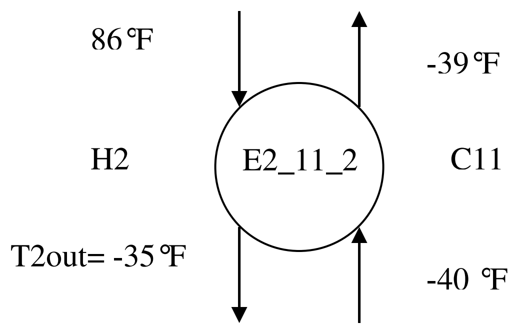

Figure B.12 Exchanger E2_11_2 calculation

For E5_9_2: The obtained loads from LINGO also are:

Q5_9_24 = 4.85 MMBtu / h

So, heat transferred from H5 to C9= 4.85 MMBtu / h

Heat balance:

4.85= 0.72* (93-T5out)

Solving for T5out = 86 ℉

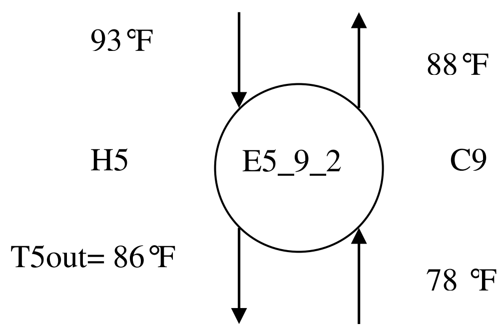

Figure B.13 Exchanger E5_9_2 calculation

For E3_13_2: The obtained loads from LINGO are:

Q3_13_39 = 13.35 MMBtu / h

So, heat transferred from H3 to C13= 13.35 MMBtu / h

Heat balance:

13.35= 4.15* (-138 -T3out)

Solving for T3out = -141 ℉

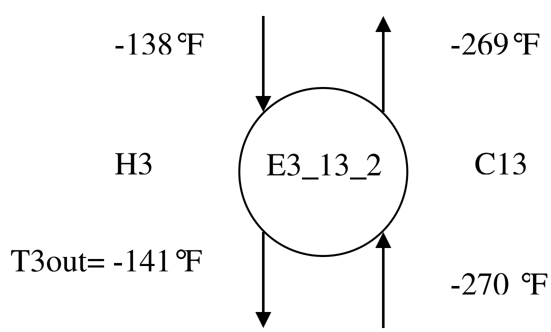

Figure B.14 Exchanger E3_13_2 calculation

For E7_13_2: The obtained loads from LINGO also are:

Q7_13_339 = 488.67 MMBtu / h

So, heat transferred from H7 to C13= 488.67 MMBtu / h

Heat balance:

488.67= 5.37* (-165 –T7out)

Solving for T7out = -265℉.

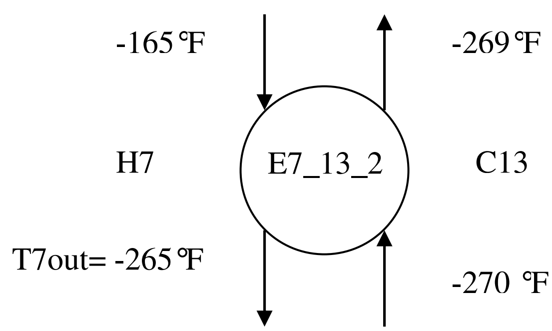

Figure B.15 Exchanger E7_13_2 calculation

Having determined the inlet and outlet temperatures of all heat exchangers, the rigorous heat transfer areas of the heat exchangers are estimated by using ASPEN Plus.

APPENDIX C

C.1 The Retrofitting Formulation Code in LINGO for Case One

```
! Objective function;
min = ((50000+1300*(A1new^0.6))*E1)+ ((50000+1300*(A2new^0.6)*E2))+
((50000+1300*(A3new^0.6)*E3))+ ((50000+1300*(A4new^0.6)*E4))+
((50000+1300*(A5new^0.6)*E5))+ ((50000+1300*(A6new^0.6)*E6))+
((50000+1300*(A7new^0.6)*E7))+ ((50000+1300*(A8new^0.6)*E8))+
((50000+1300*(A9new^0.6)*E9));

AREA_NEW=A1new+A2new+A3new+A4new+A5new+A6new+A7new+A8new+A9ne
w;
!subject to;
! Matching of Areas;
A1new+E1_1*2905+E1_2*10903+E1_3*1433+E1_4*6144+E1_5*34772+E1_6*2233+
E1_7*15076+E1_8*46484>=3973;
A2new+E2_1*2905+E2_2*10903+E2_3*1433+E2_4*6144+E2_5*34772+E2_6*2233+
E2_7*15076+E2_8*46484>=5665;
A3new+E3_1*2905+E3_2*10903+E3_3*1433+E3_4*6144+E3_5*34772+E3_6*2233+
E3_7*15076+E3_8*46484>=1939;
A4new+E4_1*2905+E4_2*10903+E4_3*1433+E4_4*6144+E4_5*34772+E4_6*2233+
E4_7*15076+E4_8*46484>=8604;
A5new+E5_1*2905+E5_2*10903+E5_3*1433+E5_4*6144+E5_5*34772+E5_6*2233+
E5_7*15076+E5_8*46484>=47778;
A6new+E6_1*2905+E6_2*10903+E6_3*1433+E6_4*6144+E6_5*34772+E6_6*2233+
E6_7*15076+E6_8*46484>=852;
A7new+E7_1*2905+E7_2*10903+E7_3*1433+E7_4*6144+E7_5*34772+E7_6*2233+
E7_7*15076+E7_8*46484>=911;
```

```
A8new+E8_1*2905+E8_2*10903+E8_3*1433+E8_4*6144+E8_5*34772+E8_6*2233+
E8_7*15076+E8_8*46484>=4081;

A9new+E9_1*2905+E9_2*10903+E9_3*1433+E9_4*6144+E9_5*34772+E9_6*2233+
E9_7*15076+E9_8*46484>=50848;

! Assignment of New Exchangers;
E1_1+E2_1+E3_1+E4_1+E5_1+E6_1+E7_1+E8_1+E9_1<=1;
E1_2+E2_2+E3_2+E4_2+E5_2+E6_2+E7_2+E8_2+E9_2<=1;
E1_3+E2_3+E3_3+E4_3+E5_3+E6_3+E7_3+E8_3+E9_3<=1;
E1_4+E2_4+E3_4+E4_4+E5_4+E6_4+E7_4+E8_4+E9_4<=1;
E1_5+E2_5+E3_5+E4_5+E5_5+E6_5+E7_5+E8_5+E9_5<=1;
E1_6+E2_6+E3_6+E4_6+E5_6+E6_6+E7_6+E8_6+E9_6<=1;
E1_7+E2_7+E3_7+E4_7+E5_7+E6_7+E7_7+E8_7+E9_7<=1;
E1_8+E2_8+E3_8+E4_8+E5_8+E6_8+E7_8+E8_8+E9_8<=1;

!Assignment of New Exchangers;

500*E1 <=A1new;
A1new<=60000*E1;
500*E2 <=A2new;
A2new<=60000*E2;
500*E3 <=A3new;
A1new<=60000*E3;
500*E4<=A4new;
A4new<=60000*E4;
500*E5 <=A5new;
A5new<=60000*E5;
500*E6 <=A6new;
```

```
A6new<=60000*E6;
500*E7 <=A7new;
A7new<=60000*E7;
500*E8 <=A8new;
A8new<=60000*E8;
500*E9 <=A9new;
A9new<=60000*E9;
@bin(E1);
@bin(E2);
@bin(E3);
@bin(E4);
@bin(E5);
@bin(E6);
@bin(E7);
@bin(E8);
@bin(E9);
@bin(E1_1);
@bin(E1_2);
@bin(E1_3);
@bin(E1_4);
@bin(E1_5);
@bin(E1_6);
@bin(E1_7);
@bin(E1_8);
@bin(E1_9);
@bin(E2_1);
@bin(E2_2);
@bin(E2_3);
@bin(E2_4);
```

```
@bin(E2_5);
@bin(E2_6);
@bin(E2_7);
@bin(E2_8);
@bin(E2_9);
@bin(E3_1);
@bin(E3_2);
@bin(E3_3);
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@bin(E3_7);
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@bin(E4_7);
@bin(E4_8);
@bin(E4_9);
@bin(E5_1);
@bin(E5_2);
@bin(E5_3);
@bin(E5_4);
@bin(E5_5);
@bin(E5_6);
```

```
@bin(E5_7);
@bin(E5_8);
@bin(E5_9);
@bin(E6_1);
@bin(E6_2);
@bin(E6_3);
@bin(E6_4);
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@bin(E6_6);
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@bin(E6_8);
@bin(E6_9);
@bin(E7_1);
@bin(E7_2);
@bin(E7_3);
@bin(E7_4);
@bin(E7_5);
@bin(E7_6);
@bin(E7_7);
@bin(E7_8);
@bin(E7_9);
@bin(E8_1);
@bin(E8_2);
@bin(E8_3);
@bin(E8_4);
@bin(E8_5);
@bin(E8_6);
@bin(E8_7);
@bin(E8_8);
```

```
@bin(E8_9);
@bin(E9_1);
@bin(E9_2);
@bin(E9_3);
@bin(E9_4);
@bin(E9_5);
@bin(E9_6);
@bin(E9_7);
@bin(E9_8);
END
```

C.2 The Retrofitting Results from LINGO for Case One

```
Global optimal solution found.
  Objective value:                              866296.7
  Extended solver steps:                             846
  Total solver iterations:                       2184697

                       Variable           Value        Reduced Cost
                          A1NEW        0.000000            0.000000
                             E1        0.000000            0.000000
                          A2NEW        5665.000            0.000000
                             E2        1.000000            232200.6
                          A3NEW        1939.000            0.000000
                             E3        0.000000            0.000000
                          A4NEW        0.000000            0.000000
                             E4        0.000000            0.000000
                          A5NEW        0.000000            0.000000
                             E5        0.000000            0.000000
                          A6NEW        852.0000            0.000000
                             E6        1.000000            74502.43
                          A7NEW        911.0000            0.000000
                             E7        1.000000            77562.96
                          A8NEW        0.000000            0.000000
                             E8        0.000000            0.000000
                          A9NEW        1000.000            0.000000
                             E9        1.000000            82026.91
                       AREA_NEW        10367.00            0.000000
                           E1_1        0.000000            148399.7
                           E1_2        0.000000            572122.2
                           E1_3        0.000000            73203.70
                           E1_4        1.000000            322399.2
                           E1_5        0.000000            1776301.
                           E1_6        0.000000            117174.1
                           E1_7        0.000000            770145.8
                           E1_8        0.000000            2439194.
                           E2_1        0.000000            76956.57
                           E2_2        0.000000            303983.1
                           E2_3        0.000000            37961.71
                           E2_4        0.000000            171298.9
                           E2_5        0.000000            921147.6
                           E2_6        0.000000            62257.56
                           E2_7        0.000000            399379.4
                           E2_8        0.000000            1296006.
                           E3_1        0.000000            148399.7
                           E3_2        0.000000            572122.2
                           E3_3        0.000000            73203.70
                           E3_4        0.000000            322399.2
                           E3_5        0.000000            1776301.
                           E3_6        0.000000            117174.1
                           E3_7        0.000000            770145.8
                           E3_8        0.000000            2439194.
                           E4_1        0.000000            148399.7
                           E4_2        1.000000            572122.2
```

E4_3	0.000000	73203.70
E4_4	0.000000	322399.2
E4_5	0.000000	1776301.
E4_6	0.000000	117174.1
E4_7	0.000000	770145.8
E4_8	0.000000	2439194.
E5_1	0.000000	148399.7
E5_2	0.000000	572122.2
E5_3	0.000000	73203.70
E5_4	0.000000	322399.2
E5_5	0.000000	1776301.
E5_6	1.000000	117174.1
E5_7	0.000000	770145.8
E5_8	1.000000	2439194.
E6_1	0.000000	-4036.809
E6_2	0.000000	0.000000
E6_3	0.000000	-1991.307
E6_4	0.000000	0.000000
E6_5	0.000000	-48319.42
E6_6	0.000000	0.000000
E6_7	0.000000	-20949.72
E6_8	0.000000	0.000000
E7_1	0.000000	0.000000
E7_2	0.000000	15150.89
E7_3	0.000000	0.000000
E7_4	0.000000	8537.746
E7_5	0.000000	0.000000
E7_6	0.000000	3102.993
E7_7	0.000000	0.000000
E7_8	0.000000	64594.49
E8_1	1.000000	148399.7
E8_2	0.000000	572122.2
E8_3	1.000000	73203.70
E8_4	0.000000	322399.2
E8_5	0.000000	1776301.
E8_6	0.000000	117174.1
E8_7	0.000000	770145.8
E8_8	0.000000	2439194.
E9_1	0.000000	5434.045
E9_2	0.000000	35545.86
E9_3	0.000000	2680.546
E9_4	0.000000	20030.61
E9_5	1.000000	65043.93
E9_6	0.000000	7280.006
E9_7	1.000000	28200.92
E9_8	0.000000	151546.7
E1_9	0.000000	0.000000
E2_9	0.000000	0.000000
E3_9	0.000000	0.000000
E4_9	0.000000	0.000000
E5_9	0.000000	0.000000
E6_9	0.000000	0.000000
E7_9	0.000000	0.000000
E8_9	0.000000	0.000000

C.3The Retrofitting Formulation Code in LINGO for Case Two

```
! Objective function;
min =(1300*(A1new^0.6)*E1)+ (1300*(A2new^0.6)*E2)+ (1300*(A3new^0.6)*E3)
+ (1300*(A4new^0.6)*E4) + (1300*(A5new^0.6)*E5) + (1300*(A6new^0.6)*E6) +
(1300*(A7new^0.6)*E7) + (1300*(A8new^0.6)*E8) + (1300*(A9new^0.6)*E9);

AREA_NEW=A1new+A2new+A3new+A4new+A5new+A6new+A7new+A8new+A9ne
w;
!subject to;

! Matching of Areas;
A1new+E1_1*2905+E1_2*10903+E1_3*1433+E1_4*6144+E1_5*34772+E1_6*2233+
E1_7*15076+E1_8*46484>=3973;
A2new+E2_1*2905+E2_2*10903+E2_3*1433+E2_4*6144+E2_5*34772+E2_6*2233+
E2_7*15076+E2_8*46484>=5665;
A3new+E3_1*2905+E3_2*10903+E3_3*1433+E3_4*6144+E3_5*34772+E3_6*2233+
E3_7*15076+E3_8*46484>=1939;
A4new+E4_1*2905+E4_2*10903+E4_3*1433+E4_4*6144+E4_5*34772+E4_6*2233+
E4_7*15076+E4_8*46484>=8604;
A5new+E5_1*2905+E5_2*10903+E5_3*1433+E5_4*6144+E5_5*34772+E5_6*2233+
E5_7*15076+E5_8*46484>=47778;
A6new+E6_1*2905+E6_2*10903+E6_3*1433+E6_4*6144+E6_5*34772+E6_6*2233+
E6_7*15076+E6_8*46484>=852;
A7new+E7_1*2905+E7_2*10903+E7_3*1433+E7_4*6144+E7_5*34772+E7_6*2233+
E7_7*15076+E7_8*46484>=911;
A8new+E8_1*2905+E8_2*10903+E8_3*1433+E8_4*6144+E8_5*34772+E8_6*2233+
E8_7*15076+E8_8*46484>=4081;
A9new+E9_1*2905+E9_2*10903+E9_3*1433+E9_4*6144+E9_5*34772+E9_6*2233+
E9_7*15076+E9_8*46484>=50848;
```

```
! Assignment of New Exchangers;

E1_1+E2_1+E3_1+E4_1+E5_1+E6_1+E7_1+E8_1+E9_1<=1;
E1_2+E2_2+E3_2+E4_2+E5_2+E6_2+E7_2+E8_2+E9_2<=1;
E1_3+E2_3+E3_3+E4_3+E5_3+E6_3+E7_3+E8_3+E9_3<=1;
E1_4+E2_4+E3_4+E4_4+E5_4+E6_4+E7_4+E8_4+E9_4<=1;
E1_5+E2_5+E3_5+E4_5+E5_5+E6_5+E7_5+E8_5+E9_5<=1;
E1_6+E2_6+E3_6+E4_6+E5_6+E6_6+E7_6+E8_6+E9_6<=1;
E1_7+E2_7+E3_7+E4_7+E5_7+E6_7+E7_7+E8_7+E9_7<=1;
E1_8+E2_8+E3_8+E4_8+E5_8+E6_8+E7_8+E8_8+E9_8<=1;

!Assignment of New Exchangers;

500*E1 <=A1new;
A1new<=60000*E1;
500*E2 <=A2new;
A2new<=60000*E2;
500*E3 <=A3new;
A1new<=60000*E3;
500*E4<=A4new;
A4new<=60000*E4;
500*E5 <=A5new;
A5new<=60000*E5;
500*E6 <=A6new;
A6new<=60000*E6;
500*E7 <=A7new;
A7new<=60000*E7;
```

```
500*E8 <=A8new;
A8new<=60000*E8;
500*E9 <=A9new;
A9new<=60000*E9;
! binary integer for exchanger retrofitting;

@bin(E1);
@bin(E2);
@bin(E3);
@bin(E4);
@bin(E5);
@bin(E6);
@bin(E7);
@bin(E8);
@bin(E9);
@bin(E1_1);
@bin(E1_2);
@bin(E1_3);
@bin(E1_4);
@bin(E1_5);
@bin(E1_6);
@bin(E1_7);
@bin(E1_8);
@bin(E1_9);
@bin(E2_1);
@bin(E2_2);
@bin(E2_3);
@bin(E2_4);
@bin(E2_5);
```

```
@bin(E2_6);
@bin(E2_7);
@bin(E2_8);
@bin(E2_9);
@bin(E3_1);
@bin(E3_2);
@bin(E3_3);
@bin(E3_4);
@bin(E3_5);
@bin(E3_6);
@bin(E3_7);
@bin(E3_8);
@bin(E3_9);
@bin(E4_1);
@bin(E4_2);
@bin(E4_3);
@bin(E4_4);
@bin(E4_5);
@bin(E4_6);
@bin(E4_7);
@bin(E4_8);
@bin(E4_9);
@bin(E5_1);
@bin(E5_2);
@bin(E5_3);
@bin(E5_4);
@bin(E5_5);
@bin(E5_6);
@bin(E5_7);
```

```
@bin(E5_8);
@bin(E5_9);
@bin(E6_1);
@bin(E6_2);
@bin(E6_3);
@bin(E6_4);
@bin(E6_5);
@bin(E6_6);
@bin(E6_7);
@bin(E6_8);
@bin(E6_9);
@bin(E7_1);
@bin(E7_2);
@bin(E7_3);
@bin(E7_4);
@bin(E7_5);
@bin(E7_6);
@bin(E7_7);
@bin(E7_8);
@bin(E7_9);
@bin(E8_1);
@bin(E8_2);
@bin(E8_3);
@bin(E8_4);
@bin(E8_5);
@bin(E8_6);
@bin(E8_7);
@bin(E8_8);
@bin(E8_9);
```

```
@bin(E9_1);
@bin(E9_2);
@bin(E9_3);
@bin(E9_4);
@bin(E9_5);
@bin(E9_6);
@bin(E9_7);
@bin(E9_8);
END
```

C.4 The Retrofitting Results from LINGO for Case Two

```
Global optimal solution found.
  Objective value:                              520561.6
  Extended solver steps:                           29947
  Total solver iterations:                       4392542
```

Variable	Value	Reduced Cost
A1NEW	0.000000	0.6828972E+09
A2NEW	0.000000	0.6828972E+09
A3NEW	0.000000	0.6828972E+09
A4NEW	0.000000	0.6828972E+09
A5NEW	1294.000	0.000000
A6NEW	852.0000	0.000000
A7NEW	911.0000	0.000000
A8NEW	4081.000	0.000000
A9NEW	1000.000	0.000000
AREA_NEW	8138.000	0.000000
E1_1	1.000000	148399.8
E1_2	0.000000	572122.8
E1_3	1.000000	75195.08
E1_4	0.000000	322399.6
E1_5	0.000000	1824622.
E1_6	0.000000	114071.2
E1_7	0.000000	770146.6
E1_8	0.000000	2374602.
E2_1	0.000000	148399.8
E2_2	0.000000	572122.8
E2_3	0.000000	75195.08
E2_4	1.000000	322399.6
E2_5	0.000000	1824622.
E2_6	0.000000	114071.2
E2_7	0.000000	770146.6
E2_8	0.000000	2374602.
E3_1	0.000000	148399.8
E3_2	0.000000	572122.8
E3_3	0.000000	75195.08
E3_4	0.000000	322399.6
E3_5	0.000000	1824622.
E3_6	1.000000	114071.2
E3_7	0.000000	770146.6
E3_8	0.000000	2374602.
E4_1	0.000000	148399.8
E4_2	1.000000	572122.8
E4_3	0.000000	75195.08
E4_4	0.000000	322399.6
E4_5	0.000000	1824622.
E4_6	0.000000	114071.2
E4_7	0.000000	770146.6
E4_8	0.000000	2374602.
E5_1	0.000000	19438.11
E5_2	0.000000	88105.72
E5_3	0.000000	11579.89

E5_4	0.000000	49648.86
E5_5	0.000000	280988.0
E5_6	0.000000	14941.59
E5_7	0.000000	100877.5
E5_8	1.000000	311036.6
E6_1	0.000000	-4036.813
E6_2	0.000000	0.000000
E6_3	0.000000	0.000000
E6_4	0.000000	0.000000
E6_5	0.000000	0.000000
E6_6	0.000000	-3102.996
E6_7	0.000000	-20949.74
E6_8	0.000000	-64594.56
E7_1	0.000000	0.000000
E7_2	0.000000	15150.90
E7_3	0.000000	1991.309
E7_4	0.000000	8537.754
E7_5	0.000000	48319.46
E7_6	0.000000	0.000000
E7_7	0.000000	0.000000
E7_8	0.000000	0.000000
E8_1	0.000000	66941.76
E8_2	0.000000	266395.6
E8_3	0.000000	35012.84
E8_4	0.000000	150117.8
E8_5	0.000000	849592.7
E8_6	0.000000	51456.44
E8_7	0.000000	347405.8
E8_8	0.000000	1071160.
E9_1	0.000000	5434.051
E9_2	0.000000	35545.89
E9_3	0.000000	4671.858
E9_4	0.000000	20030.63
E9_5	1.000000	113363.5
E9_6	0.000000	4177.017
E9_7	1.000000	28200.95
E9_8	0.000000	86952.30
E1	0.000000	0.000000
E2	0.000000	0.000000
E3	0.000000	0.000000
E4	0.000000	0.000000
E5	1.000000	0.000000
E6	1.000000	0.000000
E7	1.000000	0.000000
E8	1.000000	0.000000
E9	1.000000	0.000000
E1_9	0.000000	0.000000
E2_9	0.000000	0.000000
E3_9	0.000000	0.000000
E4_9	0.000000	0.000000
E5_9	0.000000	0.000000
E6_9	0.000000	0.000000
E7_9	0.000000	0.000000
E8_9	0.000000	0.000000

C.5 The Retrofitting Formulation Code in LINGO for Case Three

```
! Objective function;

min = A1new+A2new+A3new+A4new+A5new+A6new+A7new+A8new+A9new;

!subject to;

! Matching of Areas;

A1new+E1_1*2905+E1_2*10903+E1_3*1433+E1_4*6144+E1_5*34772+E1_6*2233+
E1_7*15076+E1_8*46484>=3973;
A2new+E2_1*2905+E2_2*10903+E2_3*1433+E2_4*6144+E2_5*34772+E2_6*2233+
E2_7*15076+E2_8*46484>=5665;
A3new+E3_1*2905+E3_2*10903+E3_3*1433+E3_4*6144+E3_5*34772+E3_6*2233+
E3_7*15076+E3_8*46484>=1939;
A4new+E4_1*2905+E4_2*10903+E4_3*1433+E4_4*6144+E4_5*34772+E4_6*2233+
E4_7*15076+E4_8*46484>=8604;
A5new+E5_1*2905+E5_2*10903+E5_3*1433+E5_4*6144+E5_5*34772+E5_6*2233+
E5_7*15076+E5_8*46484>=47778;
A6new+E6_1*2905+E6_2*10903+E6_3*1433+E6_4*6144+E6_5*34772+E6_6*2233+
E6_7*15076+E6_8*46484>=852;
A7new+E7_1*2905+E7_2*10903+E7_3*1433+E7_4*6144+E7_5*34772+E7_6*2233+
E7_7*15076+E7_8*46484>=911;
A8new+E8_1*2905+E8_2*10903+E8_3*1433+E8_4*6144+E8_5*34772+E8_6*2233+
E8_7*15076+E8_8*46484>=4081;
A9new+E9_1*2905+E9_2*10903+E9_3*1433+E9_4*6144+E9_5*34772+E9_6*2233+
E9_7*15076+E9_8*46484>=50848;

! Assignment of New Exchangers;
```

```
E1_1+E2_1+E3_1+E4_1+E5_1+E6_1+E7_1+E8_1+E9_1<=1;
E1_2+E2_2+E3_2+E4_2+E5_2+E6_2+E7_2+E8_2+E9_2<=1;
E1_3+E2_3+E3_3+E4_3+E5_3+E6_3+E7_3+E8_3+E9_3<=1;
E1_4+E2_4+E3_4+E4_4+E5_4+E6_4+E7_4+E8_4+E9_4<=1;
E1_5+E2_5+E3_5+E4_5+E5_5+E6_5+E7_5+E8_5+E9_5<=1;
E1_6+E2_6+E3_6+E4_6+E5_6+E6_6+E7_6+E8_6+E9_6<=1;
E1_7+E2_7+E3_7+E4_7+E5_7+E6_7+E7_7+E8_7+E9_7<=1;
E1_8+E2_8+E3_8+E4_8+E5_8+E6_8+E7_8+E8_8+E9_8<=1;

!Assignment of New Exchangers;

500*E1 <=A1new;
A1new<=60000*E1;
500*E2 <=A2new;
A2new<=60000*E2;
500*E3 <=A3new;
A1new<=60000*E3;
500*E4<=A4new;
A4new<=60000*E4;
500*E5 <=A5new;
A5new<=60000*E5;
500*E6 <=A6new;
A6new<=60000*E6;
500*E7 <=A7new;
A7new<=60000*E7;
500*E8 <=A8new;
A8new<=60000*E8;
500*E9 <=A9new;
A9new<=60000*E9;
```

```
! binary integer for exchanger retrofitting;
@bin(E1);
@bin(E2);
@bin(E3);
@bin(E4);
@bin(E5);
@bin(E6);
@bin(E7);
@bin(E8);
@bin(E9);
@bin(E1_1);
@bin(E1_2);
@bin(E1_3);
@bin(E1_4);
@bin(E1_5);
@bin(E1_6);
@bin(E1_7);
@bin(E1_8);
@bin(E1_9);
@bin(E2_1);
@bin(E2_2);
@bin(E2_3);
@bin(E2_4);
@bin(E2_5);
@bin(E2_6);
@bin(E2_7);
@bin(E2_8);
@bin(E2_9);
@bin(E3_1);
```

```
@bin(E3_2);
@bin(E3_3);
@bin(E3_4);
@bin(E3_5);
@bin(E3_6);
@bin(E3_7);
@bin(E3_8);
@bin(E3_9);
@bin(E4_1);
@bin(E4_2);
@bin(E4_3);
@bin(E4_4);
@bin(E4_5);
@bin(E4_6);
@bin(E4_7);
@bin(E4_8);
@bin(E4_9);
@bin(E5_1);
@bin(E5_2);
@bin(E5_3);
@bin(E5_4);
@bin(E5_5);
@bin(E5_6);
@bin(E5_7);
@bin(E5_8);
@bin(E5_9);
@bin(E6_1);
@bin(E6_2);
@bin(E6_3);
```

```
@bin(E6_4);
@bin(E6_5);
@bin(E6_6);
@bin(E6_7);
@bin(E6_8);
@bin(E6_9);
@bin(E7_1);
@bin(E7_2);
@bin(E7_3);
@bin(E7_4);
@bin(E7_5);
@bin(E7_6);
@bin(E7_7);
@bin(E7_8);
@bin(E7_9);
@bin(E8_1);
@bin(E8_2);
@bin(E8_3);
@bin(E8_4);
@bin(E8_5);
@bin(E8_6);
@bin(E8_7);
@bin(E8_8);
@bin(E8_9);
@bin(E9_1);
@bin(E9_2);
@bin(E9_3);
@bin(E9_4);
@bin(E9_5);
```

```
@bin(E9_6);
@bin(E9_7);
@bin(E9_8);

END
```

C.6 The Retrofitting Results from LINGO for Case Three

```
Global optimal solution found.
  Objective value:                              7479.000
  Extended solver steps:                              40
  Total solver iterations:                           273


                   Variable           Value        Reduced Cost
                      A1NEW        1068.000            0.000000
                      A2NEW        0.000000            1.000000
                      A3NEW        506.0000            0.000000
                      A4NEW        0.000000            1.000000
                      A5NEW        1294.000            0.000000
                      A6NEW        852.0000            0.000000
                      A7NEW        911.0000            0.000000
                      A8NEW        1848.000            0.000000
                      A9NEW        1000.000            0.000000
                       E1_1        1.000000           -2905.000
                       E1_2        0.000000           -10903.00
                       E1_3        0.000000           -1433.000
                       E1_4        0.000000           -6144.000
                       E1_5        0.000000           -34772.00
                       E1_6        0.000000           -2233.000
                       E1_7        0.000000           -15076.00
                       E1_8        0.000000           -46484.00
                       E2_1        0.000000            0.000000
                       E2_2        0.000000            0.000000
                       E2_3        0.000000            0.000000
                       E2_4        1.000000            0.000000
                       E2_5        0.000000            0.000000
                       E2_6        0.000000            0.000000
                       E2_7        0.000000            0.000000
                       E2_8        0.000000            0.000000
                       E3_1        0.000000           -2905.000
                       E3_2        0.000000           -10903.00
                       E3_3        1.000000           -1433.000
                       E3_4        0.000000           -6144.000
```

E3_5	0.000000	-34772.00
E3_6	0.000000	-2233.000
E3_7	0.000000	-15076.00
E3_8	0.000000	-46484.00
E4_1	0.000000	0.000000
E4_2	1.000000	0.000000
E4_3	0.000000	0.000000
E4_4	0.000000	0.000000
E4_5	0.000000	0.000000
E4_6	0.000000	0.000000
E4_7	0.000000	0.000000
E4_8	0.000000	0.000000
E5_1	0.000000	-2905.000
E5_2	0.000000	-10903.00
E5_3	0.000000	-1433.000
E5_4	0.000000	-6144.000
E5_5	0.000000	-34772.00
E5_6	0.000000	-2233.000
E5_7	0.000000	-15076.00
E5_8	1.000000	-46484.00
E6_1	0.000000	-2905.000
E6_2	0.000000	-10903.00
E6_3	0.000000	-1433.000
E6_4	0.000000	-6144.000
E6_5	0.000000	-34772.00
E6_6	0.000000	-2233.000
E6_7	0.000000	-15076.00
E6_8	0.000000	-46484.00
E7_1	0.000000	-2905.000
E7_2	0.000000	-10903.00
E7_3	0.000000	-1433.000
E7_4	0.000000	-6144.000
E7_5	0.000000	-34772.00
E7_6	0.000000	-2233.000
E7_7	0.000000	-15076.00
E7_8	0.000000	-46484.00
E8_1	0.000000	-2905.000
E8_2	0.000000	-10903.00
E8_3	0.000000	-1433.000
E8_4	0.000000	-6144.000
E8_5	0.000000	-34772.00
E8_6	1.000000	-2233.000
E8_7	0.000000	-15076.00
E8_8	0.000000	-46484.00
E9_1	0.000000	-2905.000
E9_2	0.000000	-10903.00
E9_3	0.000000	-1433.000
E9_4	0.000000	-6144.000
E9_5	1.000000	-34772.00
E9_6	0.000000	-2233.000
E9_7	1.000000	-15076.00
E9_8	0.000000	-46484.00
E1	1.000000	0.000000
E2	0.000000	0.000000

E3	1.000000	0.000000
E4	0.000000	0.000000
E5	1.000000	0.000000
E6	1.000000	0.000000
E7	1.000000	0.000000
E8	1.000000	0.000000
E9	1.000000	0.000000
E1_9	0.000000	0.000000
E2_9	0.000000	0.000000
E3_9	0.000000	0.000000
E4_9	0.000000	0.000000
E5_9	0.000000	0.000000
E6_9	0.000000	0.000000
E7_9	0.000000	0.000000
E8_9	0.000000	0.000000

APPENDIX D

D.1 ASPEN Plus Results

Heat and Material Balance Table			
Stream ID		5	LNG
From		N2-REJ	CRYOGHE
To		CRYOGHE	STORGLNG
Phase		MIXED	LIQUID
Substream: MIXED			
Mole Frac			
H2O		0.0	0.0
MDEA		0.0	0.0
H2S		8.51699E-9	8.51699E-9
CO2		1.68449E-7	1.68449E-7
HCO3-		0.0	0.0
MDEA+		0.0	0.0
CO3-2		0.0	0.0
HS-		0.0	0.0
S-2		0.0	0.0
H3O+		0.0	0.0
OH-		0.0	0.0
N2		5.23907E-3	5.23907E-3
EG		0.0	0.0
DEG		0.0	0.0
TEG		0.0	0.0
MEOH		0.0	0.0
C1		.9823687	.9823687
C2		.0123915	.0123915
C3		4.55873E-7	4.55873E-7
C4		7.5360E-14	7.5360E-14
C5		0.0	0.0
C6		0.0	0.0
C7		0.0	0.0
C8		0.0	0.0
C9		0.0	0.0
C10		0.0	0.0
BENZENE		0.0	0.0
TOLUENE		0.0	0.0
XYLENE		0.0	0.0
IC4		6.9782E-12	6.9782E-12
IC5		2.7406E-16	2.7406E-16
C2H4		0.0	0.0
C3H6-2		0.0	0.0
Total Flow	MMscfd	1116.863	1116.863
Total Flow	lb/hr	1.99636E+6	1.99636E+6
Total Flow	MMcuft/hr	1.552325	.0751984
Temperature	F	-164.6593	-256.0000
Pressure	psi	200.0000	18.53540
Vapor Frac		.9939751	0.0
Liquid Frac		6.02486E-3	1.000000
Solid Frac		0.0	0.0
Enthalpy	Btu/lbmol	-34218.59	-38197.78
Enthalpy	Btu/lb	-2101.969	-2346.401
Enthalpy	MMBtu/hr	-4196.276	-4684.250
Entropy	Btu/lbmol-R	-30.02445	-44.68628
Entropy	Btu/lb-R	-1.844332	-2.744974
Density	lbmol/cuft	.0789985	1.630772
Density	lb/cuft	1.286042	26.54784
Average MW		16.27931	16.27931
Liq Vol 60F	MMcuft/hr	.1059656	.1059656
*** ALL PHASES ***			
RVP-API	psi	1621.060	1621.060
*** VAPOR PHASE ***			
ZMX		.8042375	
VMX	cuft/min	25862.76	
Total Flow	MMscfd	1110.134	
CPCVMX		1.650918	

VITA

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